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JOINT PHS-LACSD WATER RENOVATION RESEARCH PROJECT

Franklin D. Dryden

Robert A. Taft Sanitary Engineering Center
Cincinnati, Ohio

1964

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JOINT PHS-LACSD
WATER RENOVATION RESEARCH PROJECT

Interim Report

Contract No. PH 86-63-214

July, 1963 - July, 1964

to

U. S. Department of Health, Education, and Welfare
Public Health Service
Division of Water Supply and Pollution Control
Basic and Applied Sciences Branch
Advanced Waste Treatment Research Program
Robert A. Taft Sanitary Engineering Center
4676 Columbia Parkway
Cincinnati 26, Ohio

from

The County Sanitation Districts of Los Angeles County

John D. Parkhurst - Chief Engineer & General Manager

by

Franklin D. Dryden - Project Co-ordinator

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Joint PHS-LACSD Water Renovation Research Project

INTRODUCTION

Franklin D. Dryden - Project Co-ordinator

Introduction

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INTRODUCTION

Chapter 1

GENERAL BACKGROUND

The Joint PHS-LACSD Water Renovation Research Project has been carried out under the provisions of Contract No. PH 86-63-214 between the United States of America and Los Angeles County Sanitation District No. 2 effective June 17, 1963. In June, 1964, the contract period was extended to September 30, 1964 without additional funds. The total project budget of $127,000.00 was shared equally by the two parties. The contract covered three distinct projects which have been designated: (1) Recreational Re-use, (2) Foam Separation, and (3) Cost-Quality.

The Recreational Reuse Study was conducted at the District 14 Water Pollution Control Plant in Lancaster, California. The purpose was to remove algae and algal nutrients from oxidation pond water to prepare it for use in recreational lakes to be constructed in the area. All applicable unit processes were to be evaluated for effectiveness and cost.

The Foam Separation Study was conducted at the District 21 Water Reclamation Plant in Pomona, California. The object was to evaluate at pilot scale all available concepts of foamer design for efficiency and cost of removing ABS and COD from an activated sludge plant effluent.

The Cost-Quality Study was conducted at the Whittier Narrows Water Reclamation Plant in El Monte, California. This facility was designed to demonstrate on a large scale the feasibility and economics of waste water renovation. Its unique situation and design features permit controlled comparison of alternative methods for operating activated sludge plants.

This interim report covers the 12 month period ending June 30, 1964. The three projects have been extended for another year under a separate contract. The report is divided into three parts, one for each of the three project study areas.

Almost all of the data available to June 30, 1964 are presented and discussed herein. Staffing and obtaining equipment essentially consumed the first half of the year with most data obtained after January 1, 1964.

The breakdown of the $127,000 expenditure was approximately $83,000.00 for Part One--Recreational Reuse, $35,000.00 for Part Two--Foam Separation and $9,000.00 for Part Three--Cost-Quality.

Chapter 2

SUMMARY OF FINDINGS

a. PART ONE - RECREATIONAL REUSE STUDY

Pilot and laboratory studies were conducted on the problem of removing algae and algal nutrients from oxidation pond water to renovate it for use in recreational lakes. Pilot processes given a preliminary trial include coagulation (with lime, alum, ferric sulfate and other coagulants and coagulant aids) followed by sedimentation and dissolved air flotation, and filtration through diatomaceous earth and sand filters. The best prospect appears to be for a combination of coagulation and dissolved air flotation followed by filtration through sand. The resulting water will be relatively clear, have a phosphate concentration of 1 to 3 mg/l, and regrow algae after exposure in lakes. The cost will probably be between 8 and 12 cents per thousand gallons.

Many variations and modifications of removal processes and chemicals remain to be tested in an effort to lower costs. As soon as possible, parallel 50 gpm processes using flotation, sedimentation and two kinds of filtration will be placed in continuous operation while additional studies are conducted concurrently. Regrowth basins will continuously receive a 1 to 5 gpm stream of product water and be evaluated chemically, bacterially, and biologically as to suitability for recreational lakes.

b. PART TWO - FOAM SEPARATION STUDY

A pilot foamer designed for 0.5 mgd capacity at a 5 minute detention time was operated with several modifications under a wide range of air rates and freeboards. ABS and COD removal were the primary criteria of effectiveness although the quantity of foamate flow was an important parameter. Removal of approximately 70% of the ABS and 35% of the COD was accomplished at 0.6 cu.ft. of air per gallon of water at a cost of about $14 per million gallons. COD of ABS and suspended solids origin comprised approximately 75% of the total COD removal.

Surface enrichment phenomena appeared to be absent. Removals were most efficient when all air used was admitted at the bottom of the foamer. Foamate flow is realated to freeboard depth and unit air rates per square foot of foamer floor area (expansion velocity). Higher freeboards and low expansion velocities can accomplish significant removals with minimal foamate production but require greater detention times and higher air rates in cu.ft./gal.

There are good prospects for using a foaming process for low cost tertiary treatment as a safety factor to remove occasionally excessive quantities of suspended solids from secondary effluents, and to economically remove some COD prior to a more expensive process like carbon absorption. After the conversion to more biodegradable detergents, it would be entirely feasible to add such detergents just ahead of the foamer. Any residual detergent not removed by the foamer would be decomposed in stream or soil.

It is also possible that foaming of primary effluents can be accomplished with the aid of polyelectrolytes. An improved foamer will be designed and operated on secondary effluent before these investigations are conducted.

c. PART THREE - COST-QUALITY

Regular collection and analysis of samples of raw sewage, primary effluent, secondary effluent, tertiary effluent and return activated sludge commenced in January, 1964. A combination of the step aeration and contact stabilization processes was used during all but a few weeks of the reporting period. Effluent of good quality was produced at flow rates from 10 to 16 mgd with plant loadings and air use remaining relatively constant. During three periods in which effluent quality showed slight deterioration, the methods used to restore desired quality are clearly evident in the data. However more time will be required to fully develop the relationships between cost and quality which comprise the study objective. Basic cost information on the plant has been prepared for future use.

With the exception that a daily sample for total carbon analysis will be added, the present series of analyses appears capable of producing the desired result. During the next year's operation, the process will be converted to standard rate activated sludge.

Chapter 3

CREDITS

a. Staff

The project staff included both Sanitation District and U. S. Public Health Service personnel. PHS personnel have been so designated in the following tabulation of all persons directly assigned to the project through June 30, 1964.

Name	Position	Starting Date
General Supervision		
John D. Parkhurst	Project Director	
Gerald N. McDermott	Project Officer (PHS)	
Franklin D. Dryden	Project Co-ordinator	
Part One - Recreational Reuse Study		
Gerald Stern	Project Engineer (PHS)	2/17/64
Jay G. Kremer	Asst. Project Engineer	9/30/63
Ricardo A. Tejidor	Chemist	10/1/63
Martel T. Durazo	Ass't. Chemist	5/22/64
David Hewitt	Laboratory Technician	8/1/63
William J. Parsonson	Field Technician	12/16/63 to 4/30/64
Earnest K. Freeman	Field Technician	4/1/64
Charles Dalton	Field Technician	6/15/64

Name	Position	Starting Date
Part Two - Foam Separation Study		
Malcolm L. Whitt	Project Engineer	(Part time)
John Zoltek, Jr.	Ass't. Proj. Eng. & Chief Investigator (PHS)	8/1/63 to 6/8/64
Nathaniel C. Anderson	Chemist	8/27/63
Part Three - Cost-Quality Study		
Franklin D. Dryden	Project Engineer	9/1/64
Malcolm Whitt	Ass't. Project Engineer	(Part time)
CSD Laboratory Personnel	Chemists	(As req'd.)

b. Others

Construction of pilot facilities in Lancaster and Pomona could not have been accomplished without the personnel of the Sanitation Districts' maintenance crews under the direction of Harry C. Chapman, Superintendent of Trunk Sewers and Pumping Stations and supervisors Clyde Hopkins and James S. Mikkelsen.

Direct help and constant cooperation was also afforded by the plant operators, C. B. Price - Lancaster, Arthur Chesley - Pomona, and Thomas F. Johnson, Whittier Narrows under the direction of Carl A. Nagel, Superintendent of Sewage Treatment, and Walter E. Garrison, Deputy Assistant Chief Engineer.

The CSD Director of Laboratories, Raymond M. Stewart, gave invaluable assistance in setting up the field laboratories and reviewing analytical procedures. Many analyses reported in each of the three studies were conducted by CSD laboratory personnel under his direction.

Technical review of the project and numerous helpful suggestions were provided by the staff of the Advanced Waste Treatment Research Program at the Robert A. Taft Sanitary Engineering Center in Cincinnati, Ohio, under the direction of F. M. Middleton, Chief, and David G. Stephan, Deputy Chief. Special appreciation is also extended to Dr. M. C. Palmer, PHS Algologist, who provided much useful information and advice on the nature of algae and on two visits to the recreational reuse project established the algae identification and counting ability of the laboratory and provided quality control throughout the study.

Joint PHS-LACSD Water Renovation Research Project

PART ONE

RECREATIONAL REUSE STUDY

Lancaster, California

Gerald Stern - Project Engineer
Jay G. Kremer - Assistant Project Engineer

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PART ONE - RECREATIONAL RE-USE STUDY

Lancaster, California

I - INTRODUCTION

A. General Background

The County Sanitation Districts of Los Angeles County and the U. S. Public Health Service have undertaken a joint project to prepare waste stabilization pond water for recreational uses. The project site is the District 14 Water Pollution Control Plant near Lancaster, California, which treats approximately 3 million gallons per day of waste water using primary sedimentation followed by secondary treatment in waste stabilization or oxidation ponds. (Figure 1)

B. Project Objective

The primary objective of the project is to remove sufficient algae, organic matter and nutrients from the oxidation pond water to allow its use in recreational lakes. The recreational lakes, which have been planned by the County Engineer as part of a regional park, will offer boating and fishing-for-fun, and possibly water contact sports. To be successful the process employed must be sufficiently economical to avoid over-pricing the water, and treatment must be sufficient to keep regrowth of algae within acceptable limits. A secondary objective of this project is to determine whether harvesting of algae for a poultry or livestock feed is feasible and will defray the cost of water production.

Acceptable limiting concentrations of algae and other contaminants in recreational lake waters have not been completely established. Based on studies at Santee, California, where recreational lakes using reclaimed waste waters have been successfully established, and recommendations of specialists with the U. S. Public Health Service, the following arbitrary criteria have been selected for the treated water:

(a) Algae concentrations of less than 10,000 counts per milliliter.

(b) Total phosphates less than 2 milligrams per liter.

(c) Turbidity of 10 Jackson candle standard turbidity units (S.T.U.) or less.

(d) Low concentrations of COD, BOD, and nitrogen.

(e) The presence of organisms of the Coliform group within the limits contained in PHS Drinking Water Standards.

(f) Pathogenic viruses, for which there are tests, to within acceptable limits.

(g) The prevention of objectionable algae regrowth.

(h) Items (a) through (g) are to be accomplished at a total maximum cost of $100 per million gallons (10 cents per thousand gallons).

C. Report Composition

This preliminary report summarizes the activities of the project for the period between September 26, 1963, and May 22, 1964. To facilitate reporting the results, the project has been divided into the following phases:

(a) Oxidation pond studies which include nutrient levels and measurements of algae concentrations and genera.

(b) Laboratory studies (jar tests) of primary coagulants and coagulant aids to reduce algae and algal nutrients and to guide pilot plant studies.

(c) Pilot plant sudies of the following process units:

1. A U. S. Army Portable Water Purification Unit (Erdlator) incorporating coagulation, upflow clarification and diatomaceous earth filtration.
2. The Infilco Corporation's Densopac process using high-lime-dosage coagulation, sedimentation and recarbonation for pH adjustment.
3. An air flotation unit including coagulation and dissolved air flotation.
4. Diatomaceous earth pressure filtration of effluents from the above processes and direct filtration of waste stabilization pond water.

(d) Future studies.

II - OXIDATION POND STUDIES

A. Objective

The objective of this phase was to determine the biological, chemical and physical characteristics of the pond waters and effluents by obtaining the following analytical information:

1. Total algae colony (clump) counts and relative counts of specific algae genera.
2. Algae nutrient levels including total phosphates and nitrogen compounds.
3. Chemical constituents including chemical oxygen demand, biological oxygen demand, total, suspended, and dissolved solids, alkalinity, and hardness.
4. Measurement of variation in chemical and physical characteristics over 24-hour periods including constituents listed in 3, dissolved oxygen, and free and total carbon dioxide.
5. Analyses to determine the distribution of nutrients and other contaminants in the waste stabilization ponds.

B. Description of Ponds

The eight oxidation ponds each average 30 acres in surface area and are normally operated at depths of 5 to 7 feet. The basic flow pattern through the oxidation ponds between September, 1963 and February, 1964 was consecutively through ponds 1, 2, 4, and 3. (Figue 2) For part of the year ponds 5 and 7 were added to the series. In February, as shown in Figure 2, the flow pattern was modified and stabilized by the use of discharge weirs for the first four ponds. Half the plant flow was introduced into Pond 1 and half into Pond 2. The Pond 1 effluent went directly into Pond 3 while the Pond 2 effluent continued to reach Pond 3 by way of Pond 4. This change probably increased the organic loading on Pond 3 although changes in pond efficiency related to changes in climate may have masked the effect of the change in the flow pattern.

Lower temperatures and shorter days decrease the level of bacterial and algal activity within the ponds. As a consequence during the winter, greater detention time is required to lower COD, suspended solids, and ammonia-nitrogen concentrations to summer levels. Since the study data is concentrated on the first four ponds of the series, the requirement for additional detention is reflected by increasing concentrations of the measured constituents during the winter months. As warm weather returns, the values will drop to their former levels.

The average detention time prior to Pond No. 3 is 45 days at 3 mgd. Ponds 1 through 4 are the only ones kept in operation all year. The other four ponds are filled as required for winter storage and emptied again in the summer to affect midge control. The sample designated "final effluent" was collected at several different locations depending on the number of ponds in service. Inconsistently low values for COD, ammonia-nitrogen, total phosphates and suspended solids in the final effluent compared to the four ponds reflect the changing detention times and sampling points. Pond sampling commenced in September, 1963.

C. Algae

The gross algae counts for ponds 1 and 2 are shown in Figure 3 and for ponds 3 and 4 in Figure 4. In most instances, the gross algae counts decreased with increased detention time. Fluctuations in algae counts are due to a number of factors including:

1. Changes in flow patterns.
2. Climatic conditions (cloudy or sunny days, strong prevailing winds, temperature variations, etc.)
3. Presence of Daphnia, Cyclops, or other crustacea which consume algae.
4. Sampling variations.
5. Counting variations.
6. Changes in pond loading.

The relative amounts of algae by genera in Ponds No. 1 through 4 are shown in Figure 5. The dominant genera in the ponds are Ankistrodesmus, Chlorella, Scenedesmus, Actinastrum, Micractinium, Oocystis, Golenkinia, and in the initial sampling (October-November period) Nitzschia, which was replaced in December by Euglena. In May, Euglena began to be replaced by Oocystis. In Ponds 3 and 4 the relative amounts appear to follow a similar pattern, although the gross algae counts differ. This would tend to indicate that the relative populations of the dominant algae genera within these ponds have been stabilized after approximately 30 days detention.

D. BOD

Biochemical oxygen demand (BOD) analyses of oxidation Ponds No. 1 through 4 were started in April. The test was performed with unfiltered pond water. The BOD levels in April were for Pond No. 1-78, Pond No. 2-53, Pond No. 4-37, and Pond No. 3-17 mg/liter, indicating a gradual decrease in BOD levels with increasing detention time. In May, the BOD levels appeared to be increasing and were for Pond No. 1-90, Pond No. 2-80, Pond No. 4-56, and Pond No. 3-57 mg/liter respectively.

E. Crustacea

Daphnia frequently appeared in Ponds No. 3 and 4 and in May were also found in Ponds No. 1 and 2. Their location and appearance within the ponds are erratic and unpredictable.

F. Chemical Data

From October through early March grab samples were taken weekly of primary effluent, water from Ponds No. 1 through 4, and the effluent from the last pond in use and were analyzed for total phosphates, ammonia-nitrogen, chemical oxygen demand (COD), total, suspended, and dissolved solids. From March through April, samples were taken fortnightly. Monthly average, maximum and minimum values are tabulated in Tables 1 through 3. Monthly COD and ammonia-nitrogen average values are plotted in Figures 6 and 7 respectively. Figures 8 and 9 present all weekly water analyses for oxidation Pond No. 3.

The data on oxidation Pond No. 3 water between October and February indicate that the chemical oxygen demand was approximately 200 mg/l, the total phosphates were approximately 40 mg/l, the total, dissolved, and suspended solids were approximately 600, 525 and 75 mg/l respectively, and the ammonia-nitrogen increased from about 1 mg/l to 15 mg/l. The change in pond flow pattern effected on February 1 may have caused some of the increased values for total and suspended solids and COD which occurred in March and April. Extensive turbulence in the ponds caused by strong winds or other climatic factors, such as the increase in temperatures or sunlight illumination may also be responsible. The increase in ammonia-nitrogen from 1 to about 15 mg/l over the winter months is largely the result of lessened algal activity and increased solubility of ammonia at lower temperatures. During the winter additional ponds were in operation and the ammonia concentration remained relatively low in the plant effluent.

Nitrogen compounds in the ponds are almost wholly in the organic nitrogen or ammonia-nitrogen form. Small concentrations, less than 1 mg/l each, of nitrite (NO_2^-) and nitrate ($NO_3^=$) nitrogen are found in the ponds. Approximately 85 to 90% of the organic nitrogen appears to be fixed within the algal cells and, therefore, will be removed from the treated water with the algae.

The primary effluent data presented in Tables 1 through 3 was derived from grab samples and samples taken from a beadchain automatic sampler which did not sample suspended material. An improved automatic sampler taking 24-hour composite samples of liquid and solids is now being developed.

G. 24-Hour Studies of Pond 3

Over 24-hour periods on April 16, 17 and May 7, 8, samples were collected from Pond 3 at 90 minute intervals and analyzed for temperature, free carbon dioxide, dissolved oxygen, alkalinity, total carbon dioxide and pH. (Figures 10 through 13) In addition, the April samples were composited into 4 six-hour samples and analyzed for COD, total phosphates, ammonia and total suspended and dissolved solids. (Figures 14 and 15) The results of these analyses generally reveal the normal variations expected in oxidation ponds.

Algal metabolism varies with the intensity of sunlight and duration of darkness. During daylight hours excess quantities of dissolved oxygen are produced by photosynthesis as new algal cells are formed of CO_2, nitrogen, phosphorous, and other trace elements. At night the algae cease to produce oxygen and instead exert a small oxygen demand at their relatively slow rate of endogenous respiration.

Bacteria, however, grow in relation to the available food and oxygen supply and tend to exert a higher demand for oxygen over the entire 24-hour period as organic matter is converted to CO_2, H_2O and other normal end products. Since CO_2 is produced throughout both day and night but only converted to algal cells during daylight hours, D.O. levels increase in the daytime and drop at night while CO_2 levels increase at night and decrease in the daylight. Colder temperatures decrease and higher temperatures increase the rate of activity for both algal cells and bacteria.

The weather during the April 16, 17 twenty-four hour period was partly cloudy and warm in contrast to the May 7, 8 twenty-four hour period, which was sunny but colder. Primarily because of the cloudy condition, the D.O. concentrations are lower for the April 16, 17 period than for the May 7, 8 period. The warmer water could have contributed to higher levels of algal activity and a raising of the D.O. level on April 16, 17 as it does in the summer if it had been in effect longer. The D.O. reached a maximum of 54% of saturation level during the April 16, 17 period as compared to 75% of saturation level for the May 7, 8 period. The average CO_2 value is higher for April 16, 17, the warmer cloudy day, due to the restriction of algal photosynthesis. Relative solubility is another factor which affects both the D.O. and CO_2 concentrations.

The pH, total alkalinity and total carbon dioxide (free CO_2 and Carbonate) levels do not vary to any appreciable extent over the 24-hour period nor between the two sampling periods. (Figure 11 compared to Figure 13)

The COD, total phosphate, ammonia and solids levels appear to vary to some extent over the 24-hour period (Figures 14 and 15), but it is less likely that these variations are due to actual changes in Pond No. 3 than variations due to sampling and analyses. The 24-hour sampling procedure will be conducted every two to four weeks during the course of this project.

H. Statistical Distribution and Reliability Study

On December 13, January 7 and January 16, grab samples were collected at 9 stations in Pond No. 3 to determine the statistical distribution of solids, ammonia-nitrogen, total phosphates and dissolved oxygen at various locations and depths as illustrated in Figure 16. The chemical analyses of each constituent are tabulated by station number in Tables 4 through 6.

The weather was cloudy and windy on all three days. However, on December 31 and January 16, the suspended solids data show a generally consistent decrease in concentration from shallow to deep, whereas on January 7th the average concentrations at all three depths are quite close implying more mixing and turbulence in the pond. (Table 11)

Variations in ammonia-nitrogen were slight and apparently unrelated to depth or location. Variations in dissolved solids and phosphates were greater and might have some relationship to daily or weekly variations in plant influent. However, normal mixing and the long detention period should have accomplished more uniform distribution of dissolved ions, and the variations may be primarily related to the accuracy of the sampling and analysis. Dissolved oxygen maximums and minimums were about 25% above and below the average values on both January 7 and 16, but the variations seem to be unrelated to depth or station.

It should be noted that the variations obtained in the

20 grab samples collected in Pond 3 appear to be as of the same order of magnitude as the variations obtained in the same constituents on grab samples between the various ponds.

The chemical analyses for total phosphates, amonia-nitrogen and COD discussed previously were for unfiltered samples which included algae and other suspended materials in the waste water. The concentrations of these contaminants in soluble form were also determined by filtering the pond water through laboratory filter paper (Schleicher and Schuell No. 589 medium) and measuring the percent change due to removing algae and other suspended matter. After filtering, approximately 80 to 90% of the total phosphate, virtually 100% of the amonia-nitrogen and approximately 50% of the COD remained in the filtered effluent.

III - LABORATORY STUDIES

A. Objective of Jar Tests

Laboratory jar tests were used to estimate the relative effectiveness and cost of various coagulants for removing enough algae to meet recreational re-use water quality standards. Jar test experiments were conducted by project personnel and by various representatives of chemical coagulant manufacturers. The laboratory jar tests were used to determine:

1. The most feasible primary coagulants and coagulant aids.
2. Coagulant dosages and treatment conditions required to reduce turbidities and algal nutrient concentrations to the desired levels.

3. Approximate costs of chemical coagulants necessary to obtain the desired water quality.

B. Methods and Equipment

The test apparatus used for the jar test study consisted of a 10-place bench scale stirring apparatus equipped with an automatic sampling system and a variable speed drive to adjust paddle revolutions to the desired mixing speeds.[1]

Standard turbidity unit (S.T.U.) measurements (Jackson candle) were used to determine the effectiveness of primary coagulants and coagulant aids in reducing the turbidity of water from oxidation Pond No. 3. The procedure for measuring S.T.U.'s was simplified by correlating Jackson candle turbidities of Pond No. 3 water at various dilutions with corresponding percent transmission on a Spectronic-20 photometer at a 375 millimicron wave length. The resulting curve (Figure 17) approximated a straight line and was used to convert percent transmission of treated and untreated Pond No. 3 water to an S.T.U. scale. When particularly good results were indicated by this rapid screening procedure additional chemical analyses were performed.

C. Comparison of Primary Coagulants

The results of jar-test coagulation studies on Pond No. 3 water using various concentrations of three primary coagulants are plotted in Figures 18 through 20. The coagulants tested were calcium hydroxide ($Ca(OH)_2$), aluminum sulfate ($Al_2(SO_4)_3$ or "alum", with and without pH adjustments, and ferric sulfate ($Fe_2(SO_4)_3$).

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[1] The procedure used was similar to the method described in the following paper: COHEN, Jesse M. "Improved Jar Test Procedure", Journal American Water Works Association, Vol. 49, No. 11, January, 1957, Pages 1425-1431.

The most effective results based on lowest final S.T.U. values were obtained with 420 mg/l of ferric sulfate and with 240 mg/l of aluminum sulfate at a pH of 5.3. However, the chemical costs of both processes were equally high (about $100 per million gallons); the lower concentrations of alum were offset by the excessive quantities of sulfuric acid required to adjust the high alkalinity water (about 300 mg/l as $CaCO_3$) from a pH of 8.5 down to 5.3. Aluminum sulfate without acid adjustment gave acceptable results at about half the cost and calcium hydroxide produced the poorest results at an intermediate cost. It should be noted, however, that in actual pilot plant studies, the jar test findings have not been substantiated. As yet, no acceptable results have been obtained with alum alone in the pilot plants and calcium hydroxide has worked far better than expected. Alum and acid have been almost as effective as predicted in the jar tests and at about the same cost.

Figures 18 through 32 indicate the effectiveness of primary coagulants on a percent S.T.U. remaining basis, rather than the absolute S.T.U. levels of the treated Pond No. 3 water. This procedure was used because Pond No. 3 turbidities vary over a wide range depending on turbulence, wind, temperature, cloud cover, and levels of biological activity. Since initial turbidities were not subject to control, percent removal appeared to be a more valid basis of comparison. The difference between the two procedures is illustrated in Figures 22 and 23 where the same data is plotted both ways for two dosages of aluminum sulfate. At the 168 mg/liter dosage of aluminum sulfate, which is apparently a borderline dosage level, the percent S.T.U. remaining appears best when the initial S.T.U. was highest, but the lowest absolute remaining S.T.U. was actually obtained at a low-intermediate initial S.T.U. This effect was essentially nullified when 240 mg/l of aluminum sulfate were used, although there are not enough points on the curve to be sure of its point of upturn in Figure 22. Absolute S.T.U. measurements are used as criteria for jar test effectiveness when using the same Pond No. 3 water sample. (Figure 21) An absolute S.T.U. level of 7 was observed with a 240 mg/liter dosage of aluminum sulfate at an acid regulated pH of 5.2. This value meets the water quality turbidity criteria of 10 S.T.U. proposed in the introduction.

D. Settling Times

The effect of settling time for calcium hydroxide coagulation is shown in Figure 18; a 24-hour settling period produced water of better clarity than a 20 to 30 minute settling period. The settling period effect for aluminum sulfate is shown in Figure 24 for 5 to 40 minute settling periods and in Figure 25 for 30 minute to 24-hour settling periods. After 30 minutes there is essentially little change in clarity when using aluminum sulfate as the primary coagulant.

E. Phosphate Removal

The reduction in total phosphate concentrations in Pond No. 3 water, with the addition of calcium hydroxide, aluminum sulfate and ferric sulfate primary coagulants are shown in Figures 26, 27, and 28 respectively. Approximately 400 mg/liter of calcium hydroxide (lime), 200 mg/liter of aluminum sulfate, or 250 mg/liter of ferric sulfate are required to reduce total phosphate levels to 2 mg/liter or less in the treated water. On a cost basis (Table 7) calcium hydroxide at 400 mg/liter is most effective. This, however, does not take into account the cost of reducing the pH to a satisfactory level for the recreational lakes. For calcium hydroxide it appears that a pH of at least 11.3 is required before reduction of total phosphate concentrations to the desired level of 2 mg/l or less is attained. This is approximately the same result reported by Rohlich.[2] Thirty minutes of coagulation and settling appears sufficient for calcium hydroxide precipitation of phosphate as calcium phosphate in a batch system. Any calcium phosphate not removed by settling would redissolve when the pH of the water is lowered by artificial or natural means.

F. Relative Costs

The effectiveness of the primary coagulants are compared on a cost basis in Figure 29. Aluminum sulfate without acid pH adjustment appears to be the best candidate on a cost basis. However, aluminum sulfate with acid adjustment of pH to a range of 5.0 to 5.6 and ferric sulfate gave the best S.T.U. reduction and the best treated water clarity at a chemical cost of approximately 100 dollars per million gallons.

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[2] ROHLICH, Gerald A., "Methods for the Removal of Phosphorus and Nitrogen from Sewage Plant Effluents"; International Conference on Water Pollution Research, London, September,1962.

Extensive jar test studies with coagulant aids and aluminum sulfate were conducted in the hopes of reducing the cost while maintaining effectiveness. A summary of the test results is tabulated in Table 8. A Union Carbide Co. cationic polyelectrolyte, UCAR-149, had the most effect in reducing aluminum sulfate dosages. However, the cost of this coagulant aid ranges from $1.00 to $1.65 per lb. ($1.25 per lb. was used for the cost data in Table 7) depending on the amount used. This high cost nullified the effect of reducing the aluminum sulfate dosage. Absolute S.T.U. levels of 10 or less were observed when using only 96 mg/liter of aluminum sulfate and relatively large concentrations of UCAR-149. The effect of UCAR-149 on reducing the percent S.T.U. remaining at constant aluminum sulfate dosages on treated Pond No. 3 water is shown in Figure 30.

The jar test phase of this project has been successful in meeting the objectives listed in the introduction of this section. The test results indicate the technical feasibility of attaining the water quality objectives for the recreational lakes by chemical coagulation. The chemical costs for attaining this water quality, hoever, are considered excessive. Future jar test studies will be specifically aimed at trying to reduce these chemical costs. Other primary coagulants such as calcium sulfate, sodium aluminate and combinations of the primary coagulants already tested will be investigated. In addition, other coagulant aids will be tested in a continuing effort to reduce the total chemical costs.

G. Air Flotation

A laboratory pressure cylinder was also used to determine the feasibility of removing turbidity and nutrients from Pond No. 3 water by air flotation. The laboratory tests indicated that air flotation was a promising process, particularly when coagulating with 240 mg/liter of aluminum sulfate prior to air flotation treatment. However, it was more convenient to study this process at pilot plant scale rather than in the laboratory.

IV - PILOT PLANT STUDIES

A. General

Pilot plant studies involved the investigation of various treatment processes to determine the feasibility of economically providing the quality water required for the recreational lakes.

The basic pilot plant research installation was built during January and February, 1964, after Erdlator testing had begun. The general area used for these test facilities is located to the north of the treatment plant sludge digesters and west of waste stabilization Pond No. 1. (Figures 1 and 33) The pilot plant installation was designed to be flexible and to accommodate various pilot scale or small production scale units. Units of up to 50 gpm can be tested in this installation. A central piping area contains influent and effluent piping for six major process equipment locations. Concrete slabs have been poured to accommodate four major pieces of equipment to date. The influent pump is located in the pilot area and is connected to 1600 feet of suction line from the north side of Pond No. 3. An effluent pump operating on electrode controls from the pilot area sump returns all waste and process waters over the dike into Pond No. 1.

The pilot units that have been tested to date are: (a) a U. S. Army, truck mounted, water treatment unit, called an Erdlator; (b) a complete water treatment process manufactured by the Infilco Corporation, the Densopac; (c) a one square foot air flotation unit manufactured by the Komline-Sanderson Corporation; (d) a one square foot diatomaceous earth pressure test filter supplied by the Johns-Manville Corporation; (e) a 31 square inch (0.215 sq. ft.) diatomaceous earth reversible cake filtration unit manufactured by the Granger Filter Corporation. Also some off-site tests were conducted on a wound wire diatomaceous earth pressure filter in the Hayward Filter Corporation's laboratories.

B. Erdlator

The U. S. Army mobile water treatment unit (Figure 33) has a rated capacity of 25 gallons per minute and the equipment consists of an upflow clarifier containing a central mixing and flocculating zone, a diatomaceous earth pressure filter, a lime slurry feeder, chemical feed pumps and various accessory pumps, tanks, etc. Suspended solids are removed in the upflow clarifier as the flow passes upward through a sludge blanket suspended about 12" below the water surface. A sludge withdrawal slot at that level was the intended means by which the blanket position and thickness was to be maintained, but the system was not effective since the algal floc tended to float quite easily.

Several primary coagulants and coagulant aids were used with the Erdlator including calcium hydroxide (lime), aluminum sulfate, sodium aluminate, ferric chloride and ferrous sulfate. Various combinations of these primary coagulants with coagulant aids such as montmorillonite clay and some cationic polyelectrolytes were also investigated.

The test results were almost all unsatisfactory. The algae floc particles, after coagulation, were extremely light with a tendency to float in the upflow clarification tank. Usually, a heavy cake of floating algae floc appeared on the clarifier surface in the afternoon. It is believed that the floc was buoyed to the surface by the attachment of bubbles of oxygen released from the supersaturated water. Based on this experience it was concluded that an air flotation or conventional settling tank might be more effective. The Erdlator clarifier was then modified to act like a poorly designed horizontal flow clarifier which produced significantly better effluents but still did not meet quality objectives.

The small diatomaceous earth filter included in the

Erdlator was operated a few times on both direct oxidation Pond 3 water and pond water which had been coagulated and settled. Several times, for about 20 to 30 minutes, good quality water was produced, after which a breakthrough of algae material degraded the effluent. Some problems were experienced with proper precoating of the filter septum and with clogging of the filter septums with algae.

Based on the tests conducted to date, it appears that the design features of the Erdlator do not suit the problem of algae removal, and tests with this unit have been discontinued for the present.

C. Densopac

The Densopac unit of the Infilco Corporation is a complete lime coagulation water treatment unit and consists of two vertical columns. (Figure 34) The first column, called the densator, is a water treatment package unit containing an initial rapid mix zone, a coagulation and flocculation zone, a settling zone and a sludge storage zone. The treated water leaving this unit flows upward through a pipe into the recarbonation unit. The volume of the densator is approximately 95 cubic feet (approximately 700 gallons storage capacity). The recommended flow rate was 25 gpm with approximately 20 to 30 minutes detention time in one column. The densator design features a relatively large lime dosage for high pH coagulation and rapid settling; recirculation of the high density sludge enhances the rapid settling characteristics of the floc. A solu-bridge control unit proportions the lime feed to the densator in accordance with the ratios of the conductivity of the influent water and the treated water in the tank. The lime slurry is fed through a progressing-cavity positive displacement pump which operates intermittently, in response to the controller.

The recarbonator consists of eight inverted trays where carbon dioxide comes in contact with the densator effluent to adjust the pH. Sludge accumulations are recirculated back to the inlet of the recarbonator to assist in the removal of calcium carbonate precipitate.

Initial attempts to develop a high density sludge in the Densopac were unsuccessful although a medium density sludge was developed. The organic content of the sludge inhibited rapid settling and extensive and frequent carryover of algal floc particles from the densator to the recarbonator occurred. This floc carryover persisted despite reduced process flow and sludge recirculation rates.

To prevent algae floc carryover it was found necessary to waste sludge either intermittently or continuously. Intermittent blow down of sludge to waste was used initially, but it upset the steady state operations. Beginning in early April, a slow continuous blow down of sludge to waste was instituted to stabilize the sludge blanket level.

The maximum successfully treated flow rate was approximately 15 gpm, although 10 gpm permits easier control of the unit. The overflow rate is 1 to 1.5 gpm per square foot with a detention time of 40 to 50 minutes in each unit. Although lime ($Ca(OH)_2$) is the primary coagulant used in the densator, coagulant aids were tested in both the densator with lime and in the recarbonator with carbon dioxide. Coagulant aids include Superfloc 16 (a non-ionic polymer now called Magna floc),UCAR-149 (a cationic polyelectrolyte), WPB (a non-ionic polymer), ferric sulfate, aluminum sulfate, calcium carbonate and calcium chloride. None of the aids has significantly improved the pilot test results. Calcium chloride is effective in reducing the high sodium alkalinity concentrations in the oxidation pond water, but it is more expensive than lime. At times, the addition of calcium chloride (up to 100 mg/liter) appeared to produce a fast settling floc, but this effect was difficult to maintain because of the need for continuous waste sludge blow down.

Test results are summarized in Tables 9, 10 and 15 and Figure 31. In general, coagulation at high lime dosages (over 650 mg/liter) produces an effluent that meets the turbidity objective of 10 S.T.U. and a value of 5 S.T.U. was achieved on one run.

Table 10 summarizes the alkalinity and hardness data for the Densopac tests. Table 11 summarizes the effectiveness of the Densopac in removing turbidity, phosphate, ammonia, and C.O.D. from Pond 3 water. Again, large lime dosages were needed to reduce total phosphate concentrations to a satisfactory level without filtration. Lower lime dosages (approximately 400/mg/liter) could be used provided either filtration or an extended settling period was provided to remove finely precipitated calcium phosphate. There is little or no ammonia-nitrogen removal despite the high pH of the treated water. Algae counts have been as low as 10,000/ml in the product water.

Variations in the total solids content of flow passing through the Densopac unit is largely due to the addition of lime as a coagulant. The addition of lime ($Ca(OH)_2$) to the Densator influent flow, which contains much bicarbonate (HCO_3^-) or carbon dioxide (CO_2) alkalinity, causes the calcium to go into solution forming calcium bicarbonate ($Ca(HCO_3)_2$) and increasing the total and dissolved solids. The cations sodium, potassium and ammonia, which were originally combined with the bicarbonate anion, form soluble hydroxides and do not precipitate as shown below:

$2(Na+,\ K+ \text{ or } NH_4+)HCO_3^- + Ca(OH)_2 \longrightarrow Ca(HCO_3)_2 + 2(Na+,\ K+$ or $NH_4+)OH$. As more lime is added the calcium bicarbonate tends to be converted to the slightly soluble calcium carbonate as shown below:

$$Ca(HCO_3)_2 + Ca(OH)_2 \longrightarrow 2CaCO_3 + 2H_2O$$

Any more lime added then goes into solution as calcium hydroxide or remains as suspended particles. The relative solubilities of calcium carbonate, calcium hydroxide and calcium bicarbonate are in the ratios 1 to 1.64×10^3 to 8.8×10^8 respectively. In the recarbonator unit carbon dioxide, CO_2, is added to the flow containing dissolved and suspended calcium hydroxide. This addition causes the less soluble calcium carbonate to precipitate as shown below:

$$Ca(OH)_2 + CO_2 \longrightarrow CaCO_3 + H_2O$$

This reaction decreases the total dissolved solids through reduction of calcium hydroxide and concurrent precipitation of minor amounts of other compounds such as calcium phosphate. Thus, as the flow passes through the Densopac unit, an initial build up in total and dissolved solids occurs in the densator as calcium hydroxide is added. As flow proceeds through the recarbonator, carbon dioxide is added and the total and dissolved solids are reduced through calcium carbonate precipitation and other concurrent precipitations.

The test results indicate that it is technically feasible to produce water suitable for use in recreational lakes by the Densopac process. However, the cost of the water exceeds the cost objectives for the study, even though it may still be competitive with imported supplies.

The chemical coagulant costs for producing low turbidity water are estimated to be $100.00 per million gallons excluding the considerable cost of carbon dioxide for pH adjustment, if it cannot be allowed to occur naturally. Excessive lime dosages were required to obtain water of desired clarity (95% removal of turbidity). If cost objectives are going to be met, it may be necessary to raise the turbidity (S.T.U.) objectives for the treated water. Densopac treatment with approximately 400 mg/liter of lime has produced water that appears to meet all quality objectives, except for clarity. Supplemental filtration, planned for extensive testing in the near future, will improve the clarity and general quality of the treated water. Whether the increased cost of filtering will offset the savings on chemical coagulants remains to be seen.

D. Air Flotation

The pilot air flotation unit had a one square foot surface area and was manufactured by the Komline-Sanderson Corp. (Figure 35) The rated capacity of this unit was 0.3 to 3.0 gpm at air pressures of 60 to 70 psi; the majority of tests were conducted at one gpm. Algae laden water from Pond 3 was coagulated and fed by gravity to the bottom of the flotation unit where it mixed with the high pressure fresh water supply supersaturated with dissolved air. In a full size unit, recycled effluent would be used for the pressured stream, instead of the fresh water. Flocculation facilities were designed and built to add and mix coagulants prior to air flotation treatment. The coagulation and flocculation tanks were constructed of two 55 gallon oil drums with fabricated paddle stirrers that could be adjusted to control mixing speed. One barrel provided a flocculation detention time of about 40 minutes. Initial laboratory air flotation experiments indicated the need for coagulation prior to air flotation. Primary coagulants and coagulant aids tested were aluminum sulfate, calcium hydroxide, ferric sulfate, ferrous sulfate, sulfuric acid and UCAR-149 cationic polyelectrolyte.

Aluminum sulfate dosages ranged from 120 to 240 mg/liter, although earlier experiments had indicated that over 200 mg/l would be required to produce a satisfactory reduction in turbidity. Based on laboratory experience, the pH was regulated to about 5.5 with sulfuric acid at an estimated cost of $42.00 per million gallons. Results obtained with aluminum sulfate coagulation are shown in Table 12. The best results were obtained with aluminum sulfate dosages of approximately 240 mg/liter at a pH of approximately 5.5. Both the S.T.U. levels and total phosphate concentrations were reduced substantially. (Run Number 5-Table 12) Lower aluminum sulfate dosages produced effluent of poorer clarity and increased total phosphate concentrations.

Poor results were observed with aluminum dosages of 240 mg/liter without pH adjustment. (Runs 6 and 7-Table 12) These poor test results are believed due to break-up of the aluminum sulfate-algae floc as it passes through the influent piping to the air flotation unit. The piping of a larger size air flotation unit should eliminate this floc break-up, and test runs using aluminum sulfate without pH regulation are planned.

The float skimmed from the surface of the air flotation unit was generally frothy, like fine-bubble detergent foam, light to medium in consistency, and light to dark green in color. (Figure 35) Attempts were made to harvest the float material. The frothy condition of the float made drying of the cake by natural means somewhat difficult because of the water trapped in the algal floc. The dried algae tended to form thin gray brittle flakes.

Chemical costs for aluminum sulfate coagulation were about $100.00 per million gallons or greater. (Table 12) These costs partially reflect the sulfuric acid pH adjustment which was greater than anticipated because of the high alkalinity in Pond 3 water. The cost of UCAR-149 also proved to be substantial.

Calcium hydroxide coagulation of algae followed by air flotation gave significantly better results than previously indicated in the laboratory. Lime dosages ranged from 200 to 800 mg/liter. Good quality effluents were produced with lime dosages of 400 mg/liter to 800 mg/liter. (Table 13 and Figure 32) Optimum algae coagulation with lime appeared to be influenced by the pH levels attained during treatment. A pH of 11.3 or above was essential for best results. This observation was also noted in the Densopac test operations. The pH of 11.3 or higher is also required to remove substantial amounts of calcium phosphate. Lime added to the air flotation unit was premixed with fresh water in a 0.2 to 0.4 pounds per gallon slurry. This lime preparation process may have contributed to the improvement in results obtained on the pilot air flotation unit compared to those obtained in the laboratory.

In Table 14 the solids data before and after treatment by aluminum sulfate coagulation and air flotation are summarized.

Harvesting of the float produced by lime flocculation was more successful than on the aluminum sulfate floc float. The material was dense, contained 6% solids, dried easily in the sun, and produced small, light-weight cakes of algae and lime. A partial chemical analysis of the cake material indicated the following composition:

1. Total phosphates - 17%
2. Calcium carbonate - 19%
3. Total Kjeldahl nitrogen (organic nitrogen plus ammonia-nitrogen) - 5.5%
4. Chemical oxygen demand - 10%
5. Dry weight - 97%

The test results support the finding that air flotation, like the Densopac, is technically capable of producing water acceptable for use in recreational lakes.

Testing of the Komline-Sanderson air flotation unit has been discontinued in favor of a 50 gpm Pacific Industrial Engineering unit placed in operation in June. Future tests will emphasize the production of suitable quality water while reducing chemical costs. The new air flotation unit, plus a 50 gpm diatomaceous earth filter, now on order, should provide a basis for evaluating the cost of a combined air flotation-filtration process.

E. Diatomaceous Earth Filtration

Diatomaceous earth pressure filtration was investigated using a one square foot filter obtained from the Johns-Manville Corporation for approximately 6 days in April on effluent from the Densopac recarbonator, effluent from the air flotation unit and untreated Pond 3 water.

Filtration of effluents from the air flotation and recarbonator units produced treated water with S.T.U. levels of about 5 or less and algae counts of zero to 20,000 per milliliter. The filter cycles ranged from 2.5 to 7 hours. Both pre-coat of the filter septum and body feed with diatomaceous earth (D.E.) were necessary to produce good quality water. Diatomaceous earth costs are estimated to be approximately $60 to $70 per million gallons. A larger D.E. filter, better operating controls, and possibly recovery and re-use of D.E. could reduce these costs.

Filtration of Pond 3 water directly, without previous treatment, resulted in effluents with an S.T.U. level of about 85 on an influent of 170 S.T.U., which amounts to approximately a 40 to 50 percent reduction in turbidity. Length of filter run was only about 2.5 hours, and filtration costs are estimated at nearly $80 per million gallons.

The backwash cake from the direct filtration of pond water separated easily into a settled D.E. layer and a concentrated suspension of algae which was further concentrated by centrifuging and drying. Based on this small scale measurement, it is estimated that approximately 500 pounds of algae could be harvested per million gallons of Pond 3 water. The D.E. could possibly be recycled after separating it from the algae backwash and reconditioning it in a furnace; in this way D.E. costs might be reduced by 80 to 85 percent (based on information in the literature). It is estimated that D.E. filtration would cost approximately $25 to $30 per million gallons with D.E. recovery and recycle. The market value of the harvested algae is also estimated to be about $25 per million gallons. Filtration economics apparently depends upon the feasibility and efficiency of separating and re-using the diatomaceous earth.

The Granger reversible cake filtration unit has been operated briefly, chiefly for algae harvesting and for direct filtration of Pond 3 water. This filter uses a short reversible cycle which permits a low head loss through the filter cake. Separation of diatomaceous earth from the backwash solution is possible by sedimentation or centrifugation. Only D.E. precoat is necessary for filtration because of the short cycle time of 1 to 3 minutes. Test results thus far have shown that the filter enhances the possibility of harvesting algae directly from Pond 3 water, but that the filter effluent has an S.T.U. of approximately 85 or similar to that obtained in the pressure D.E. filter. This unit produced water of approximately 25 S.T.U. when 50% D.E. and 50% nuchar activated carbon (by volume) were used as precoat before direct filtration of Pond 3 water. Additional studies with this unit are planned.

Off-site tests of a diatomaceous earth pressure filter were conducted in the laboratory of the Hayward Filter Corp. The Hayward filter consists of a central perforated tube on which a wound wire membrane is fitted. On the two available tubes, the space between adjacent wire windings is 12 and 30 microns. According to the manufacturer, the filter requires less diatomaceous earth for precoat and body feed. The filter has been operated on both raw oxidation pond water and on air flotation unit effluent. Tests to date have indicated the need to add substantial amounts of D.E. or other filter aid materials to obtain satisfactory effluents. Additional on-site experimentation is planned for this unit.

F. Summary of Pilot Studies

Judged by the accomplishment of arbitrary quality objectives, coagulation of oxidation pond water followed by either air flotation or settling will produce an effluent suitable for use in recreational lakes. Excessively expensive concentrations of chemical coagulants have been required to achieve these results and only after operating holding basins for extended periods will the true suitability of the product water be verified.

Lime appears to be the most promising primary coagulant although the cost and need for pH reduction of lime treated waters have yet to be evaluated. Production of carbon dioxide from digester gas may not materially reduce the cost of the process and certainly could not compete with adsorbing carbon dioxide from the atmosphere if the latter can be accomplished in the requisite time. The use of aluminum sulfate without pH adjustment may surpass lime treatment if the laboratory results can ever be duplicated on pilot scale. However, if algae harvesting for use as a poultry feed is ever practised, lime would be a more desirable coagulant than aluminum sulfate.

Although similar results have been obtained with both the air flotation and Densopac units, there is more room for improvement of the flotation process than the Densopac, which has been tested more extensively and received the technical assistance of several Infilco engineers. Filtration will most likely be required and definitely will improve the effluent quality, but there is still much to be done to determine the range of cost and applicability for diatomaceous earth, sand, and anthracite filtration.

V - FUTURE STUDIES

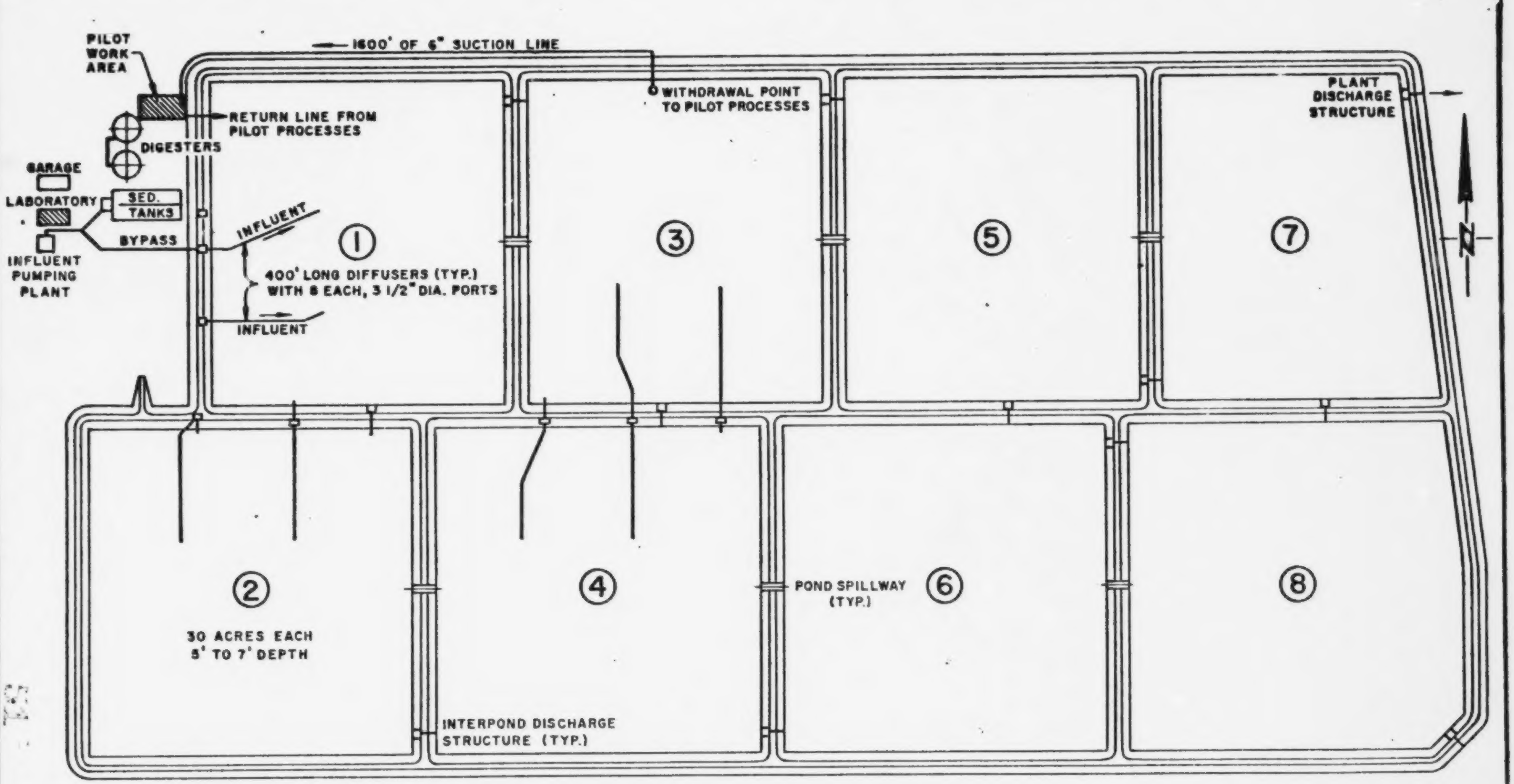
During fiscal year 1964-65, major emphasis will be placed on establishing and lowering process costs while testing product quality in simulated lakes or holding basins. Operation of the Densopac and air flotation units will be placed on a 24-hour basis for extended periods to confirm that previous results can be both duplicated and sustained. It is anticipated that some type of filtration will be an essential part of the process, and all types of filters will be studied on full pilot scale.

Although the original concept of the study was to develop a removal process capable of working on oxidation pond water as found in uncontrolled ponds, it is possible that the economics of algae removal will be improved if forced growth of selected strains of algae can produce dense algal blooms. Since North American Aviation is already involved in this phase of the problem at the plant site, it is anticipated that the basic removal processes now being developed will be applied to their high-rate pond effluents.

The holding basins now under development will not only reveal whether the original quality objectives were reasonable but will allow extended detention periods to assist viral and bacterial monitoring.

The laboratory will continue to give technical support to the pilot phase of the study and in addition will conduct evaluation tests on many of the potential coagulant aids still untried. It is also hoped that a gross evaluation of bacterial populations in the ponds and in the regrowth basin can be accomplished.

It is anticipated that by the end of the year the most economical process capable of producing water with the requisite characteristics will have been selected and tested on a full pilot scale.


PILOT WORK AREA
RETURN LINE FROM PILOT PROCESSES
DIGESTERS
GARAGE
LABORATORY
SED. TANKS
BYPASS
INFLUENT PUMPING PLANT
1600' OF 6" SUCTION LINE
WITHDRAWAL POINT TO PILOT PROCESSES
PLANT DISCHARGE STRUCTURE
INFLUENT
INFLUENT
400' LONG DIFFUSERS (TYP.) WITH 8 EACH, 3 1/2" DIA. PORTS
①
②
③
④
⑤
⑥
⑦
⑧
30 ACRES EACH 5' TO 7' DEPTH
POND SPILLWAY (TYP.)
INTERPOND DISCHARGE STRUCTURE (TYP.)
N

NOTE: FOR FLOW PATTERN, SEE FIG. 2

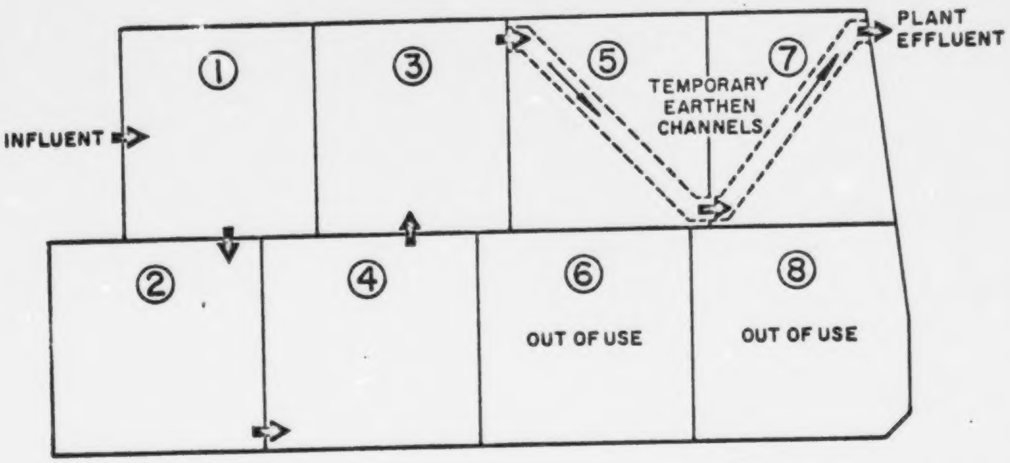
DISTRICT 14 WATER POLLUTION CONTROL PLANT

GENERAL LAYOUT OF

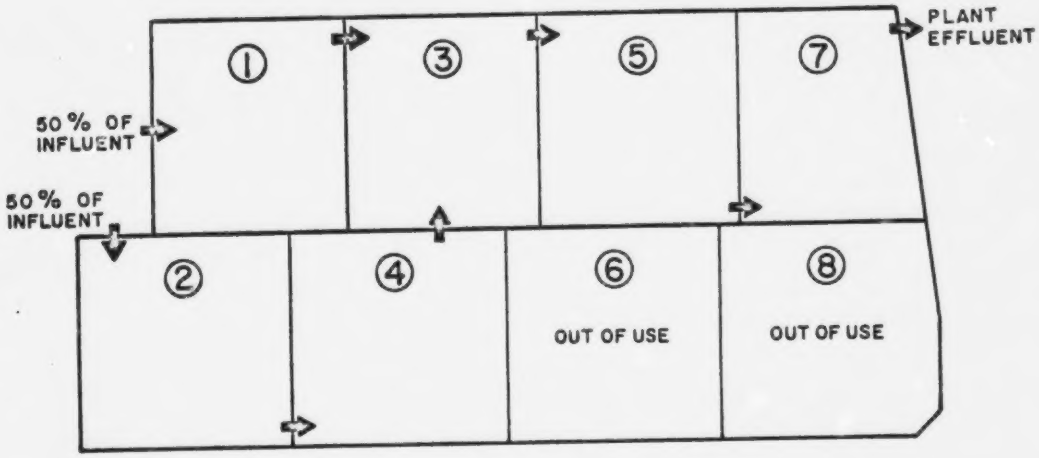
OXIDATION PONDS 1 THRU 8 AND THE STUDY AREA

SCALE 1" = 500'

FIG. 1


PLANT EFFLUENT
①
③
⑤
⑦
TEMPORARY EARTHEN CHANNELS
INFLUENT
②
④
⑥
⑧
OUT OF USE
OUT OF USE

A. POND FLOW PATTERN FROM SEPT. 1963 TO FEB. 1, 1964
FLOW BETWEEN PONDS IS CONTROLLED BY SLIDEGATES


PLANT EFFLUENT
①
③
⑤
⑦
50 % OF INFLUENT
50 % OF INFLUENT
②
④
⑥
⑧
OUT OF USE
OUT OF USE

B. POND FLOW PATTERN SINCE FEBRUARY 1, 1964
(DURING THIS PERIOD PONDS 5 & 7 WERE GRADUALLY FILLED)
ON FEB. 1, CONTROL WEIRS WERE PLACED IN THE INTERPOND STRUCTURES FOR PONDS 1 THRU 4 AND THE INFLUENT WAS DIVIDED BETWEEN PONDS 1 AND 2.

DISTRICT 14 WATER POLLUTION CONTROL PLANT
FLOW PATTERNS FOR OXIDATION PONDS 1 THRU 8

FIG. 2

ALGAE COUNTS IN PONDS NO. 1 AND NO. 2

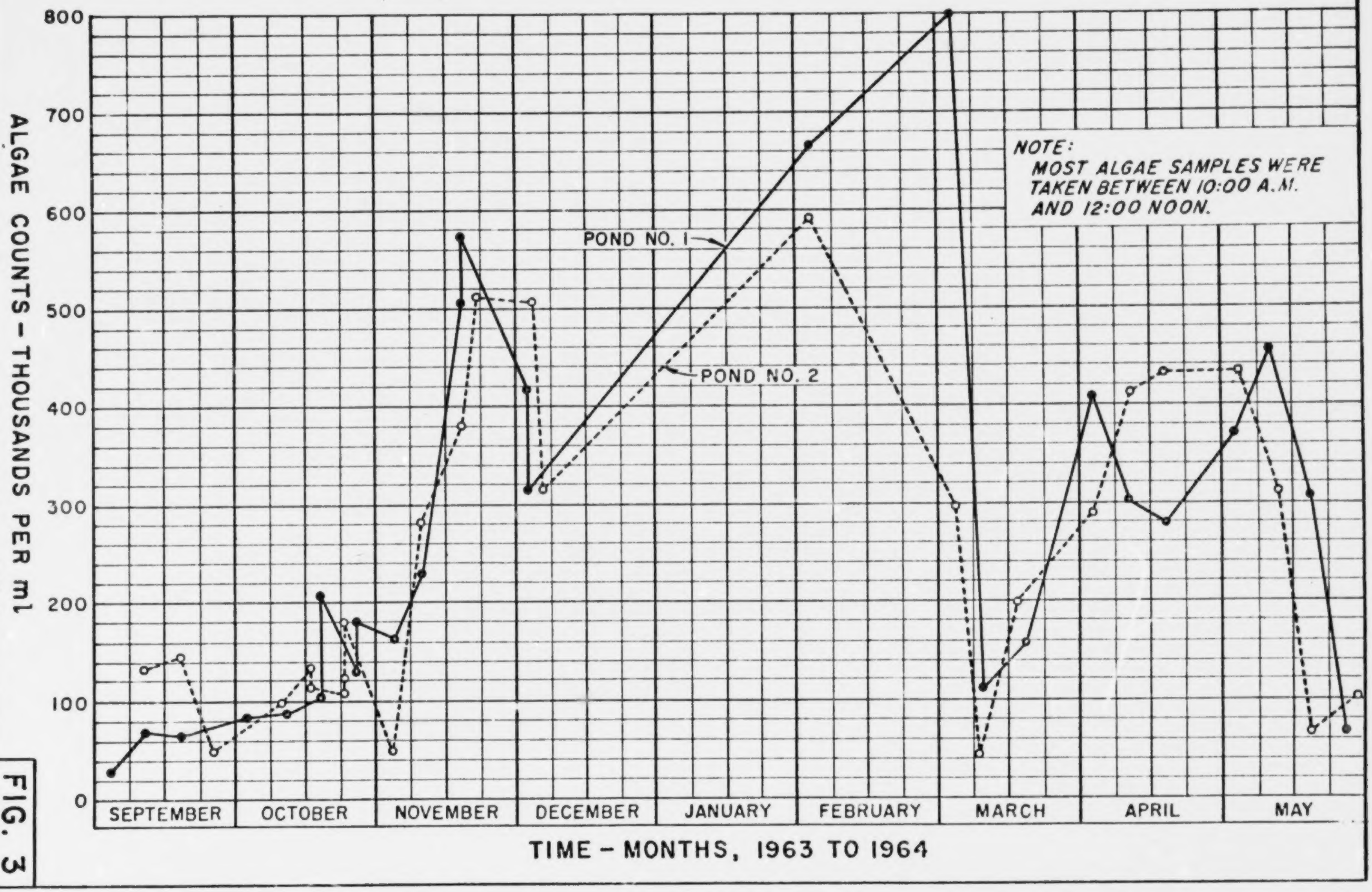


FIG. 3

ALGAE COUNTS IN PONDS NO. 3 AND NO. 4

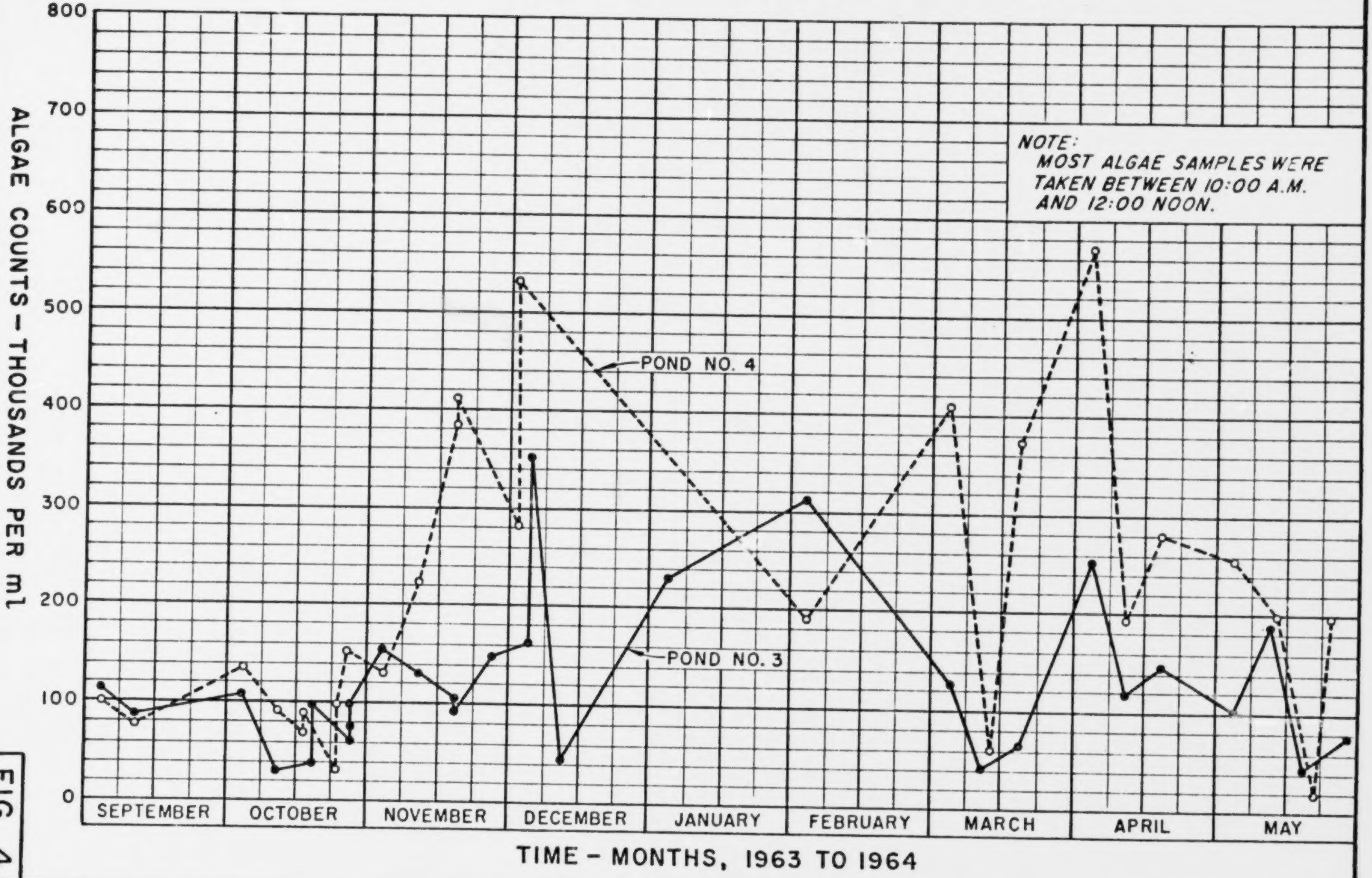


FIG. 4

Scenedesmus	April-May 1964	XXXXXXX	XXXXXXX	XXXXXX	XXXXXXX
Oocystis		XXXXXXX	XXXXXXXX	XXXXXXXXX	XXXXXXXX
Micractinium		XXXX	XX	X	XXX
Golenkinia		XX	XX	X	X
Euglena		XXXXXX	XXXXX	XXXXXXX	XXXXX
Cyclotella		XXX	XX	XX	XX
Chlorella		XXXXX	XXXX	XXXX	XXXXX
Chlamydomonas		XXX	XX	XXXX	XXX
Ankistrodesmus		XXXXXXXX	XXXXXXXXX	XXXXXXXX	XXXXXXX
Actinastrum		XXXX	X	X	X
Scenedesmus	February-March 1964	XXXXXXXX	XXXXXX	XXXXXXXX	XXXXXXXX
Oocystis		XXXXXX	XXXX	XXXXXXX	XXXXXXX
Micractinium		XXXX	X	XX	XX
Golenkinia		XXXX	XX	XX	XX
Euglena		XXXXXXXX	XXXXXXXXXX	XXXXXXXX	XXXXXXXX
Cyclotella		X	XX	XX	X
Chlorella		XXXXXX	XXXX	XXXXXX	XXXXXX
Chlamydomonas		XXXX	XXXXXX	XXXXXX	XXXXXX
Ankistrodesmus		XXXXXXXX	XXXXXXXX	XXXXXXXX	XXXXXXXX
Actinastrum		XXXXXX	XX	XX	XX
Scenedesmus	December-January 1963-1964	XXXXXXXXXX	XXXXXXXXXX	XXXXXXXX	XXXXXXXXXX
Oocystis		XXXX	XXXXX	XXXXXX	XXXXX
Micractinium		XXXXXX	XXXX	XXXXXX	XXXXXX
Golenkinia		XXXXXX	XXXXXX	XXXXXX	XXXXXX
Euglena		XXXXXX	XXXXXX	XXXXXX	XXXXXXXX
Cyclotella		XX	XX	XX	X
Chlorella		XXXXXX	XXXX	XXXX	XXXX
Chlamydomonas		XXXXXX	XXXXXXXX	XXXXXX	XXXXXX
Ankistrodesmus		XXXXXXXX	XXXXXXXX	XXXXXX	XXXXXXXX
Actinastrum		XXXXXX	XXXX	XXXXXX	XXXXXX
Scenedesmus	October-November 1964	XXXXXXXX	XXXXXXXX	XXXX	XXXXXXXX
Oocystis			XXXX	XXXX	XXXXXX
Micractinium		XXXX	XXXX	XXXX	XXXX
Golenkinia		XXXX	XXXXXX	XXXXXX	XXXX
Euglena		X	XXX	XX	XX
Cyclotella		XXXXXX	XXXXXX	XXXXXX	XXXXXX
Chlorella		XXXXXXXX	XXXXXX	XXXXXX	XXXXXX
Chlamydomonas		XXXX	XXXXXX	XXXX	XXXX
Ankistrodesmus		XXXXXX	XXXXXX	XXXX	XXXXXX
Actinastrum		XXXXXX	XXXXXX	XXXX	XXXX
Predominant Algae Genera		Relative Concentrations Plotted on a 0 to 10 Scale			
		POND 1	POND 2	POND 4	POND 3

RELATIVE AMOUNTS OF THE PREDOMINANT ALGAE GENERA
IN PONDS 1 THROUGH 8, BI-MONTHLY AVERAGES

FIGURE 5

CHEMICAL OXYGEN DEMAND IN WASTE STABILIZATION PONDS

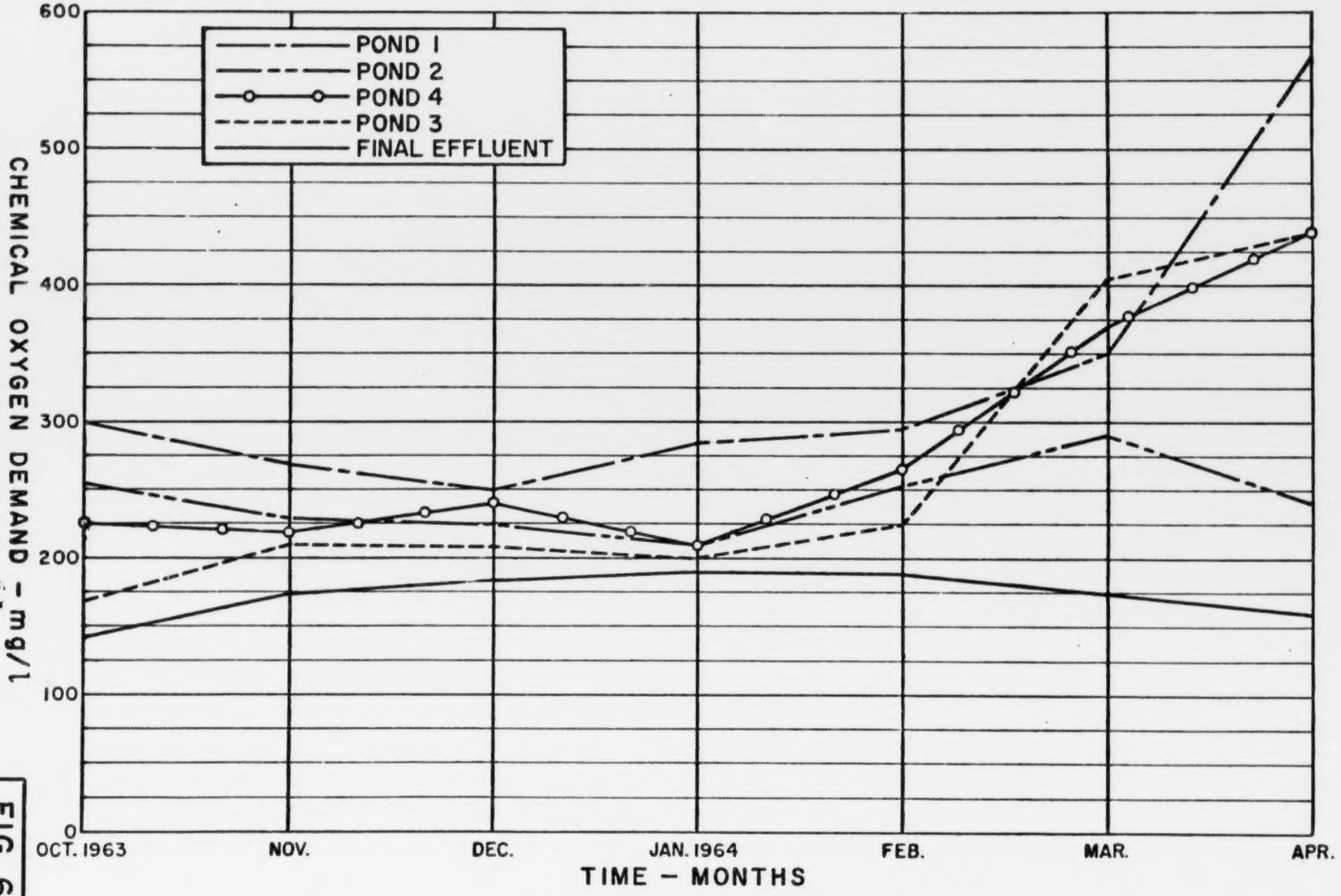


FIG. 6

56

40

AMMONIA NITROGEN IN WASTE STABILIZATION PONDS

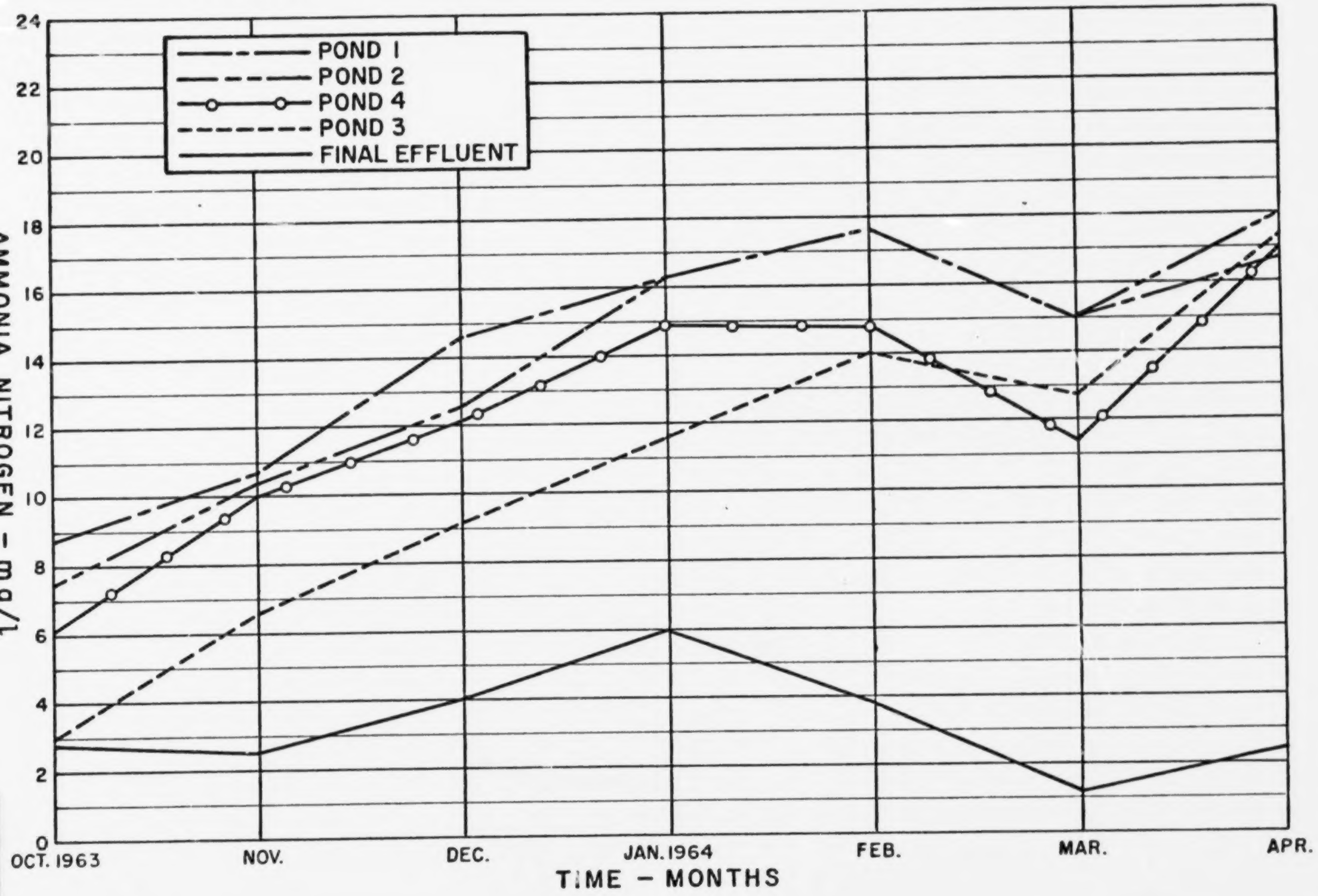


FIG. 7

CHEMICAL OXYGEN DEMAND, TOTAL PHOSPHATE AND AMMONIA NITROGEN CONCENTRATIONS IN POND 3

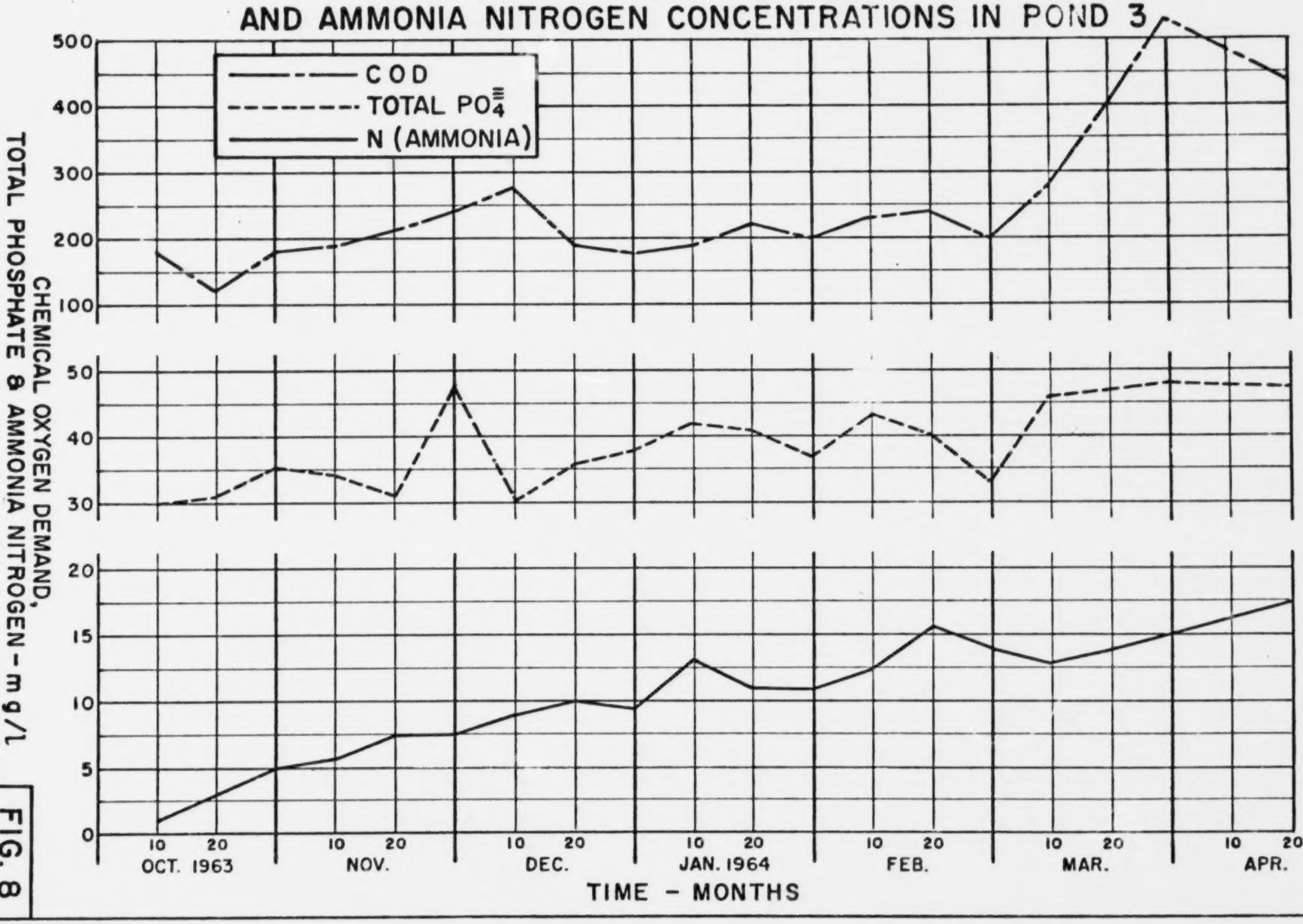


FIG. 8

TOTAL, DISSOLVED AND SUSPENDED SOLIDS CONCENTRATIONS IN POND 3

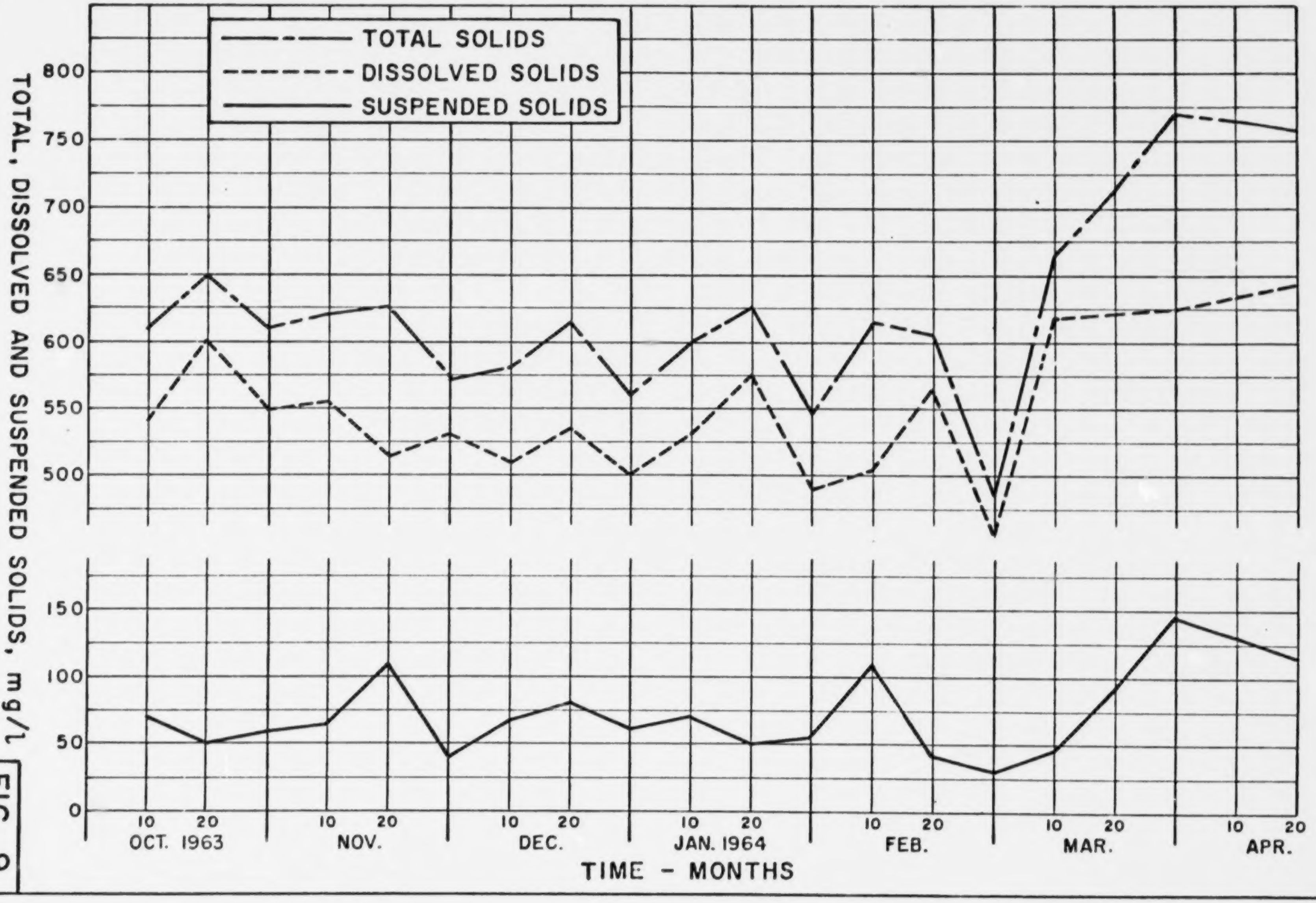


FIG. 9

POND 3 - 24 HOUR ANALYSES FOR TEMPERATURE, DISSOLVED OXYGEN AND FREE CARBON DIOXIDE

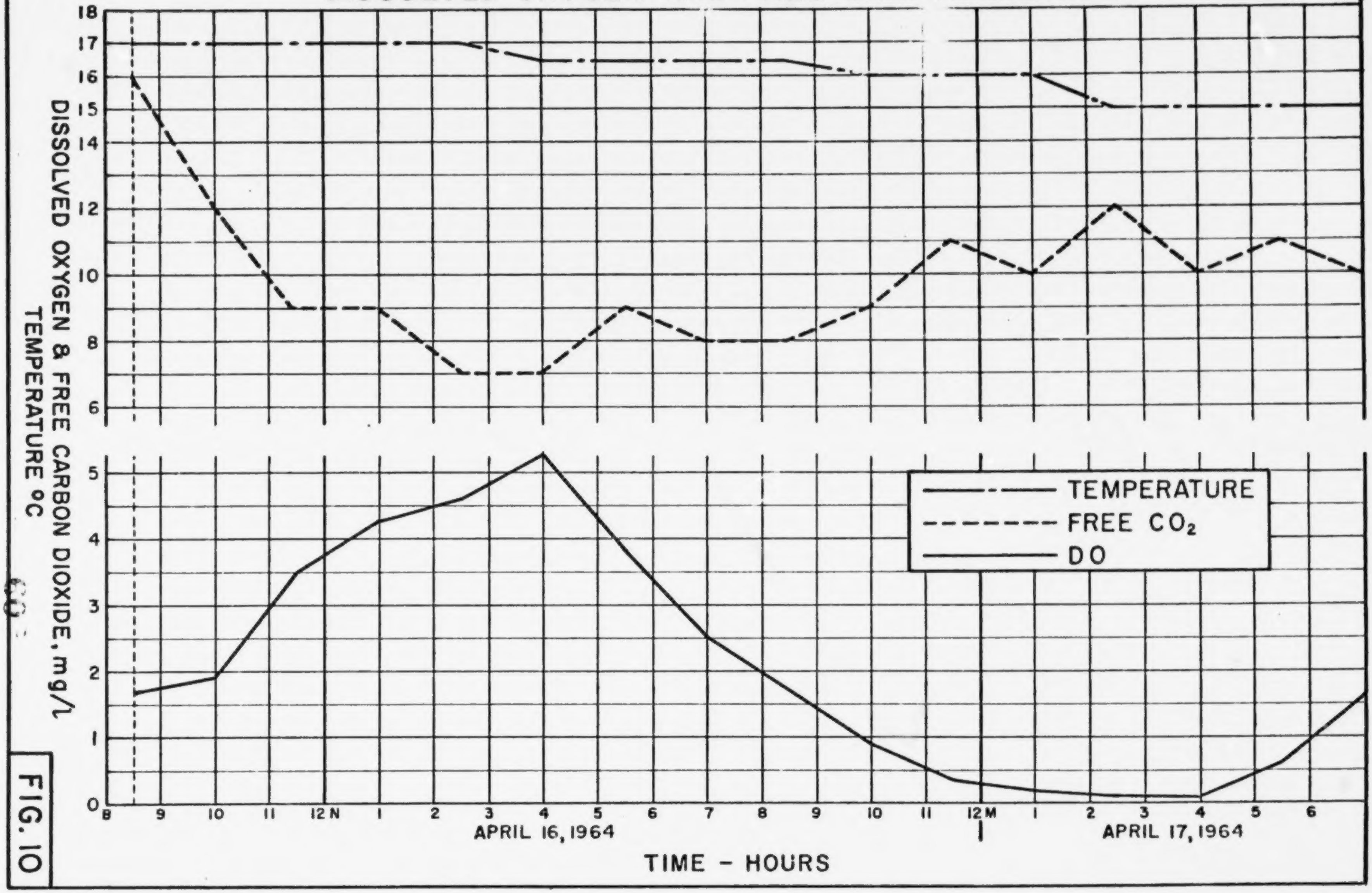
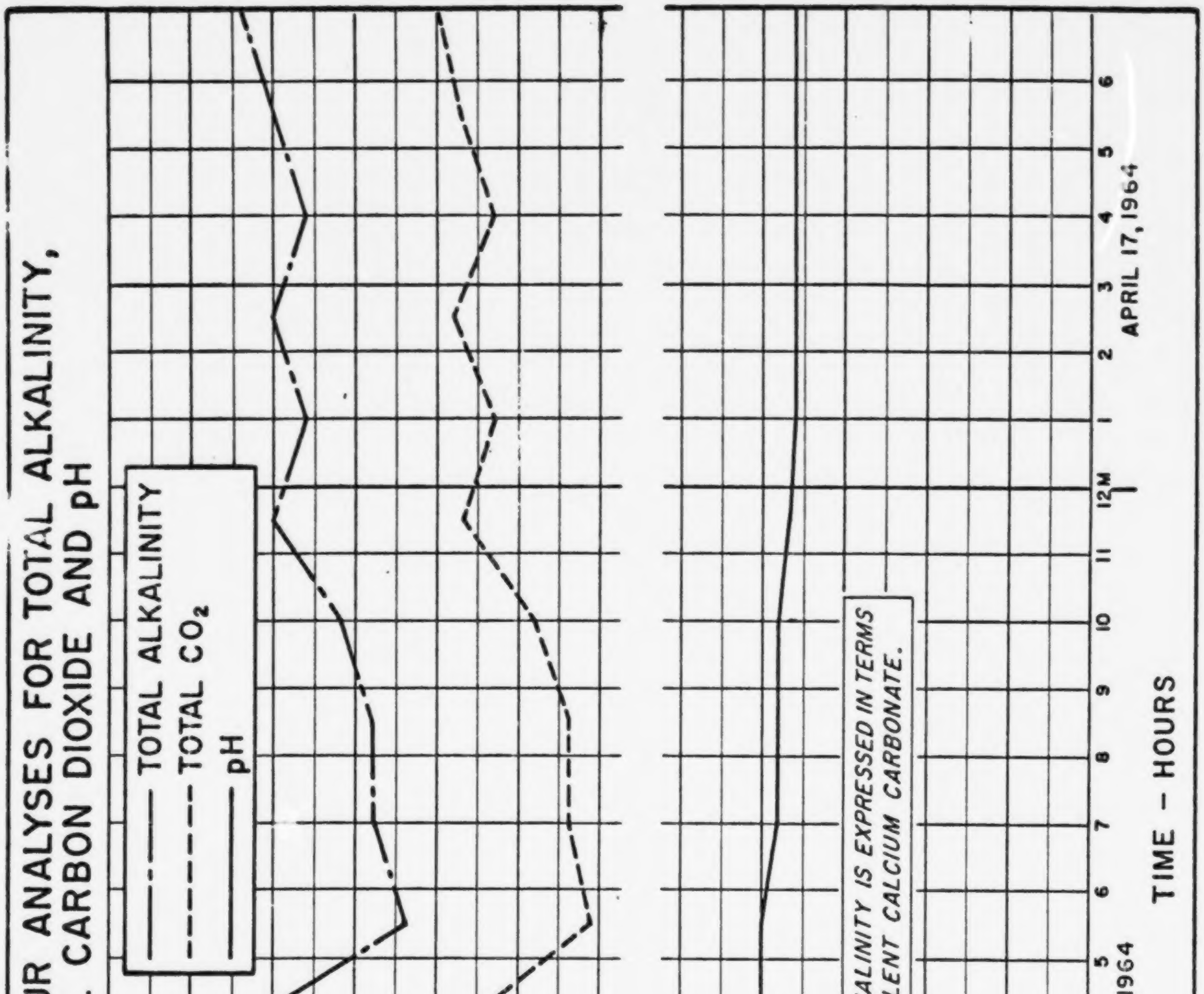


FIG. 10



POND 3 - 24 HOUR ANALYSES FOR TOTAL ALKALINITY, TOTAL CARBON DIOXIDE AND pH

NOTE: TOTAL ALKALINITY IS EXPRESSED IN TERMS OF EQUIVALENT CALCIUM CARBONATE.

FIG. 11

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POND 3 - 24 HOUR ANALYSES FOR TEMPERATURE, DISSOLVED OXYGEN AND FREE CARBON DIOXIDE

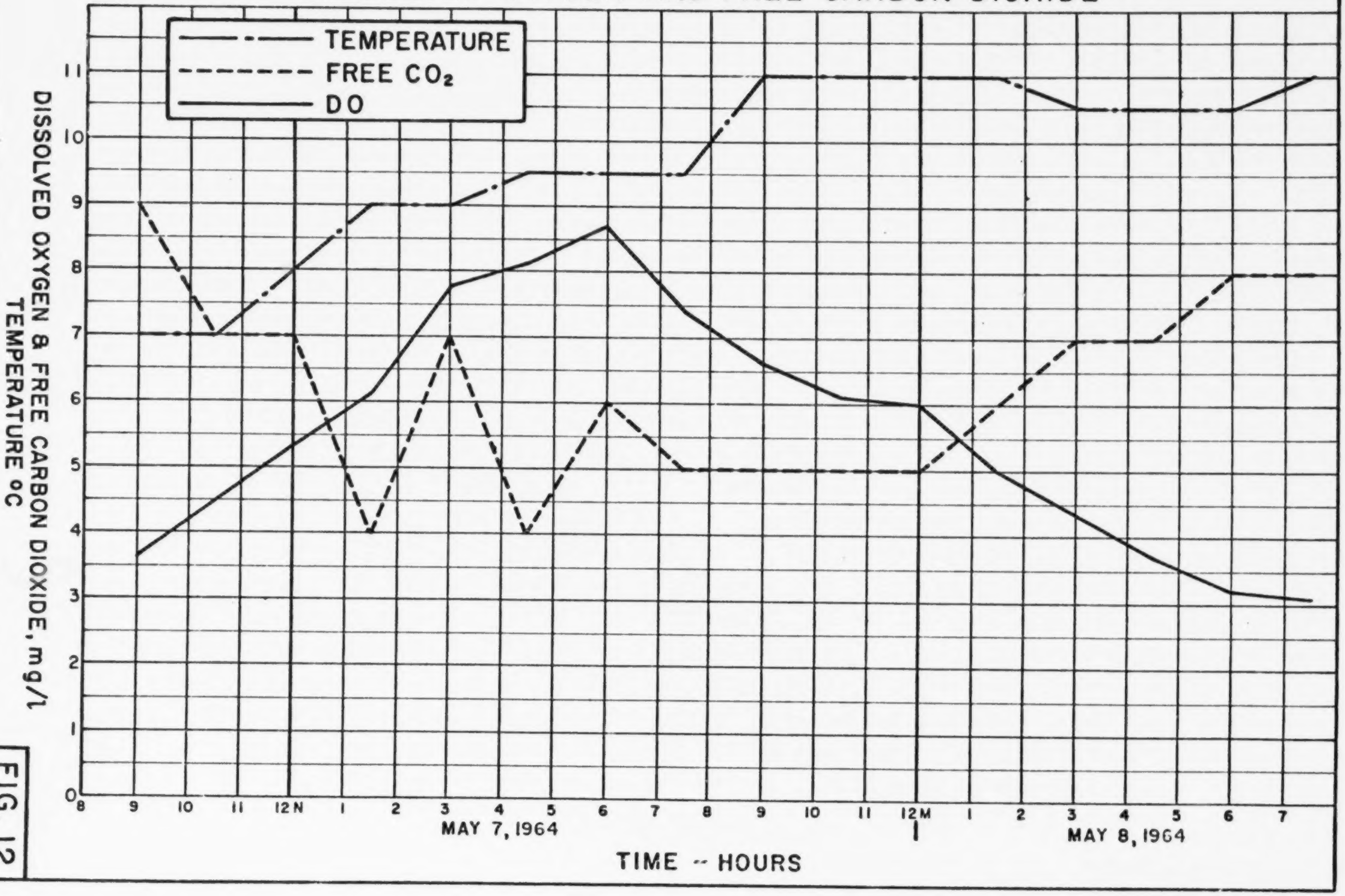
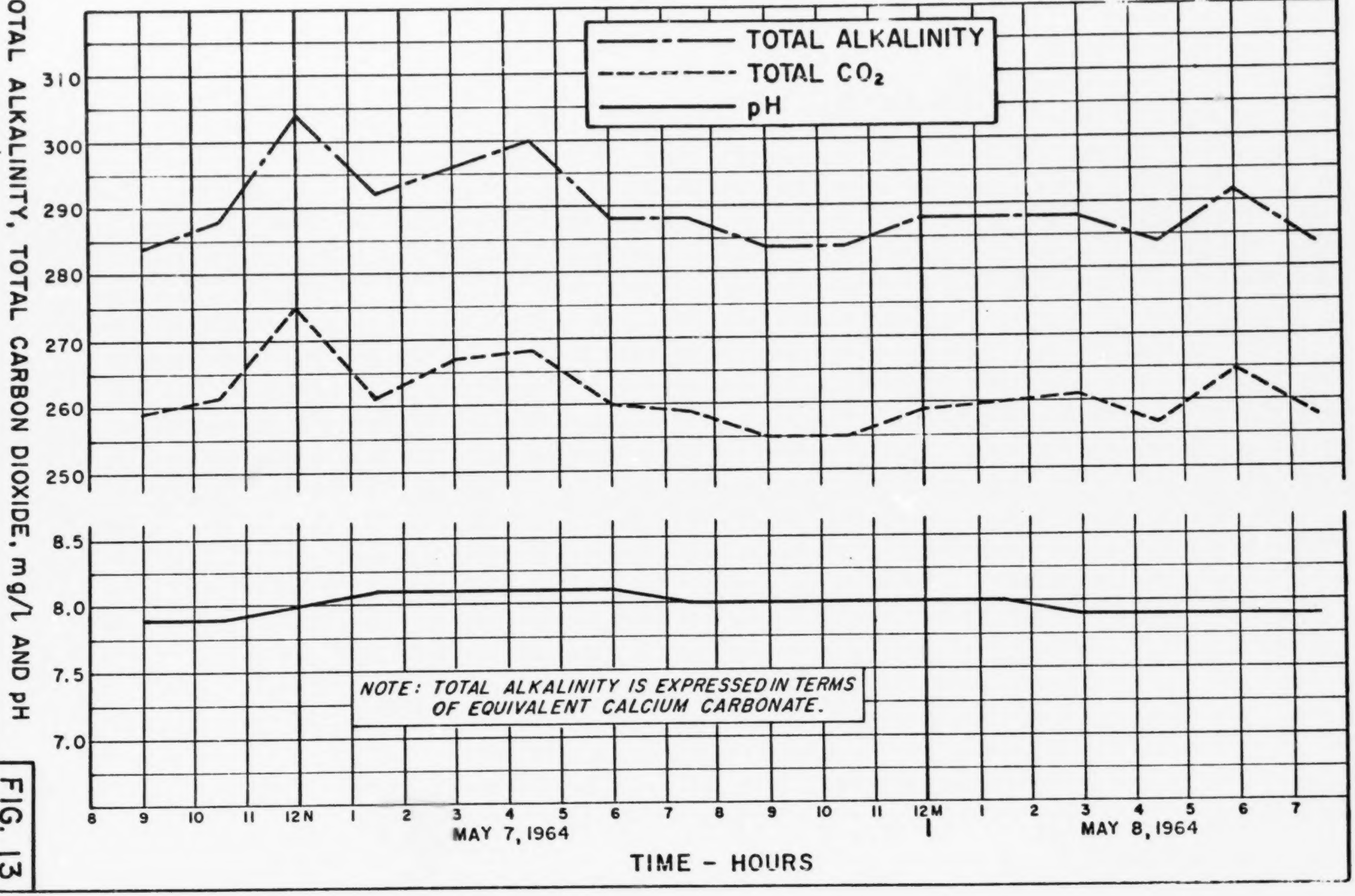


FIG. 12

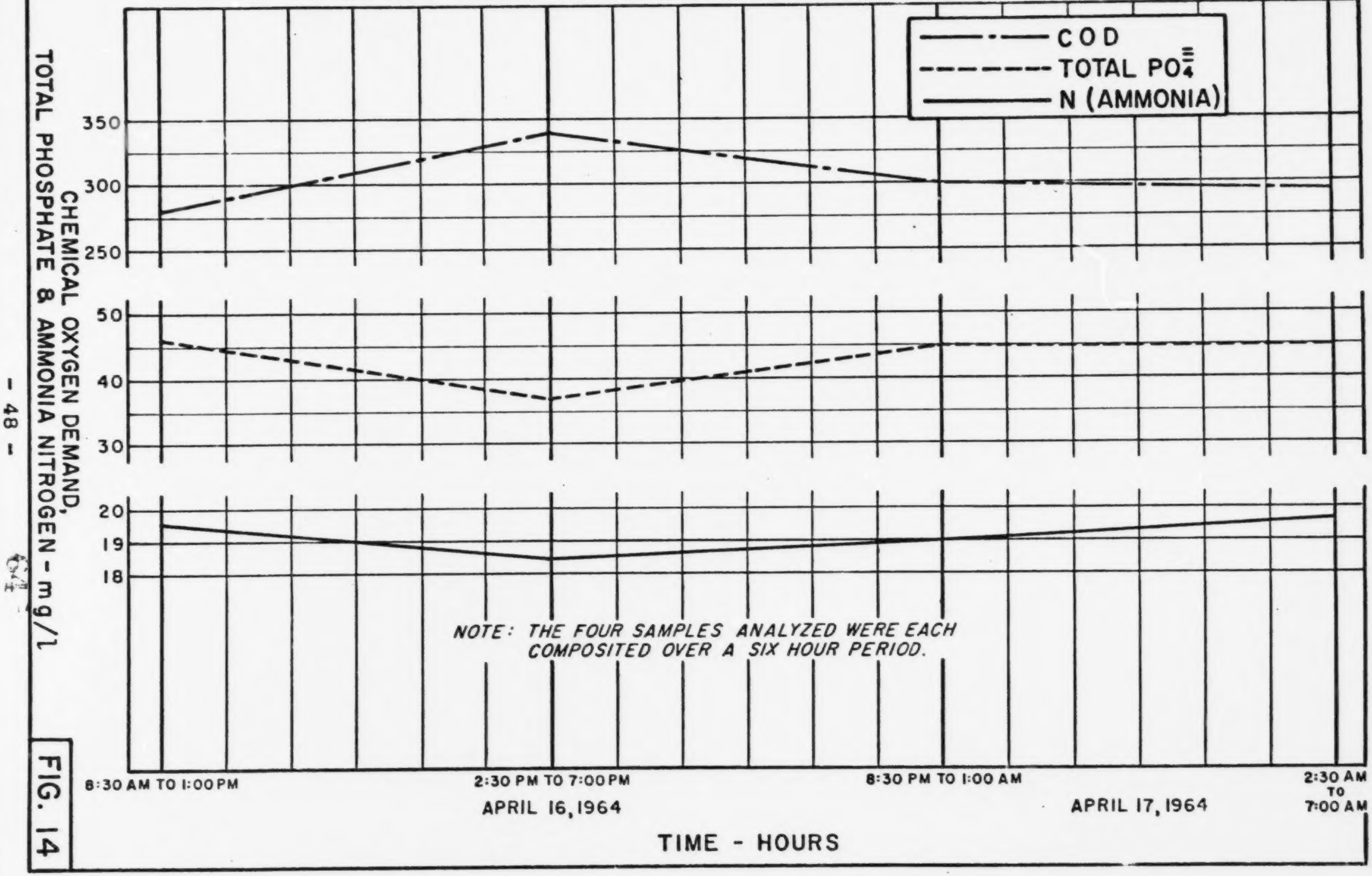
POND 3 - 24 HOUR ANALYSES FOR TOTAL ALKALINITY, TOTAL CARBON DIOXIDE AND pH



NOTE: TOTAL ALKALINITY IS EXPRESSED IN TERMS OF EQUIVALENT CALCIUM CARBONATE.

FIG. 13

POND 3 - 24 HOUR ANALYSES FOR CHEMICAL OXYGEN DEMAND, TOTAL PHOSPHATE AND AMMONIA NITROGEN



NOTE: THE FOUR SAMPLES ANALYZED WERE EACH COMPOSITED OVER A SIX HOUR PERIOD.

FIG. 14

POND 3 - 24 HOUR ANALYSES FOR TOTAL, DISSOLVED AND SUSPENDED SOLIDS

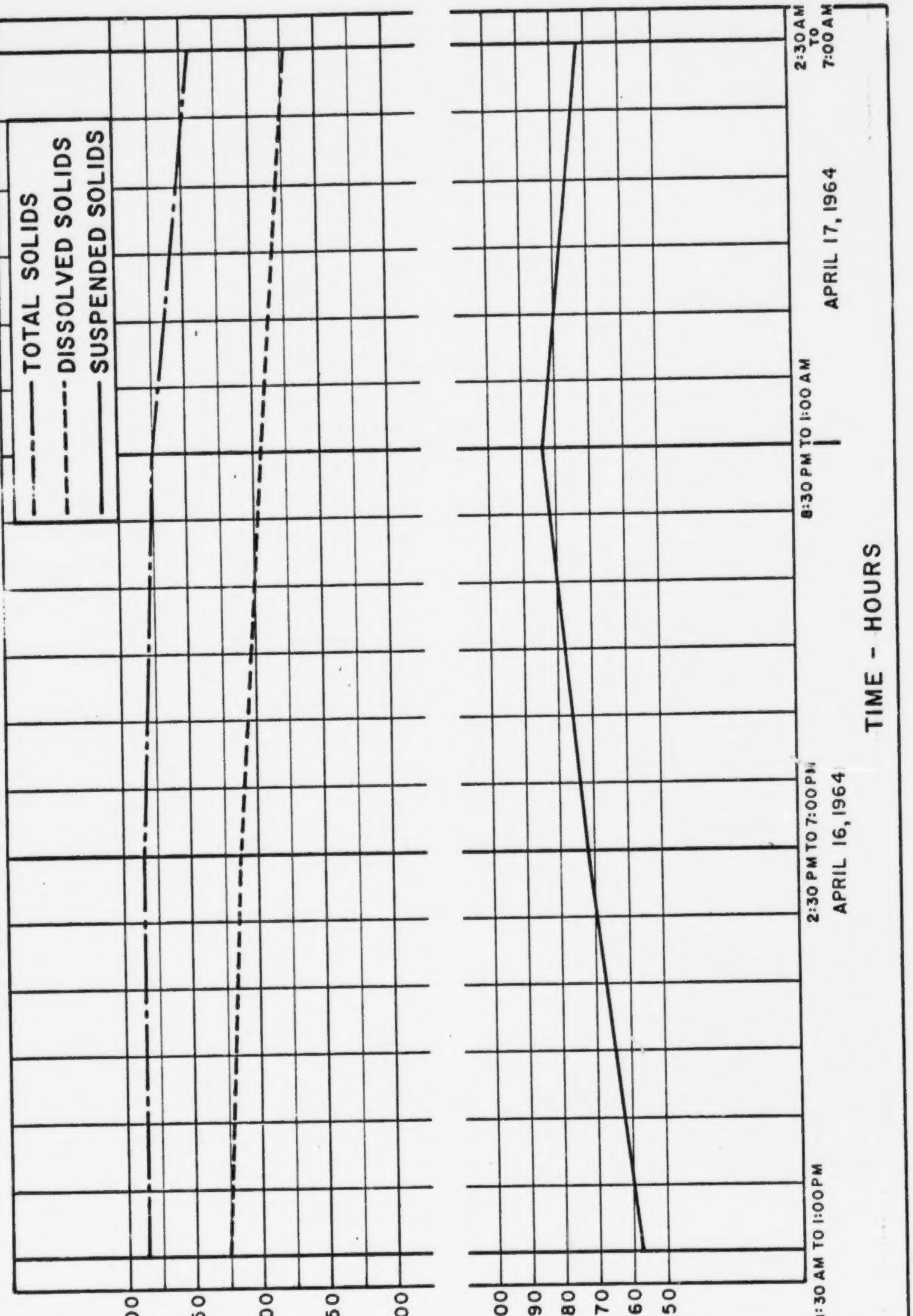
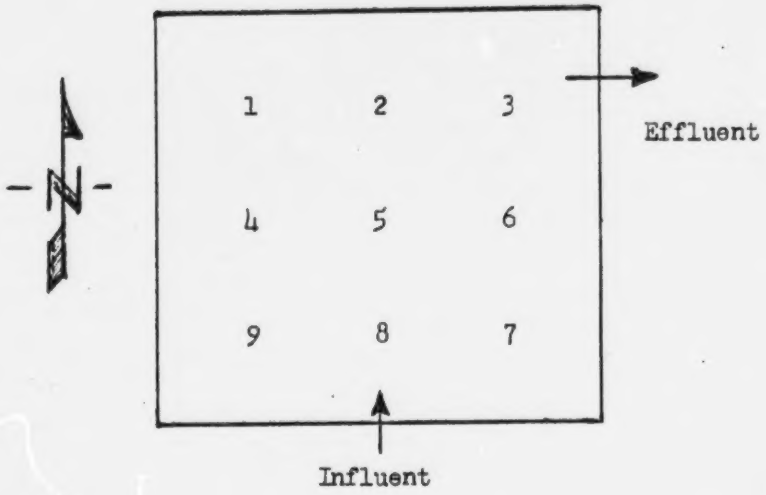


FIG. 15

65



LOCATION OF SAMPLING STATIONS IN POND 3

STATIONS SAMPLED AT EACH DEPTH

S (Surface)	M (Middle)	L (Lower)
1	-	1
-	2	2
3	3	-
4	-	4
5	5	5
-	6	6
7	-	7
8	8	-
-	9	-

Note: SURFACE samples were taken 4-inches below the surface. MIDDLE samples were taken 30-inches below the surface. LOWER samples were taken 54-inches below the surface. Influent and effluent samples were collected 30-inches below the surface adjacent to the respective structures.

DETAILS OF SAMPLING POINTS FOR

DISTRIBUTION ANALYSES OF POND NO. 3

FIGURE 16

PERCENT TRANSMITTANCE vs. STANDARD TURBIDITY UNITS

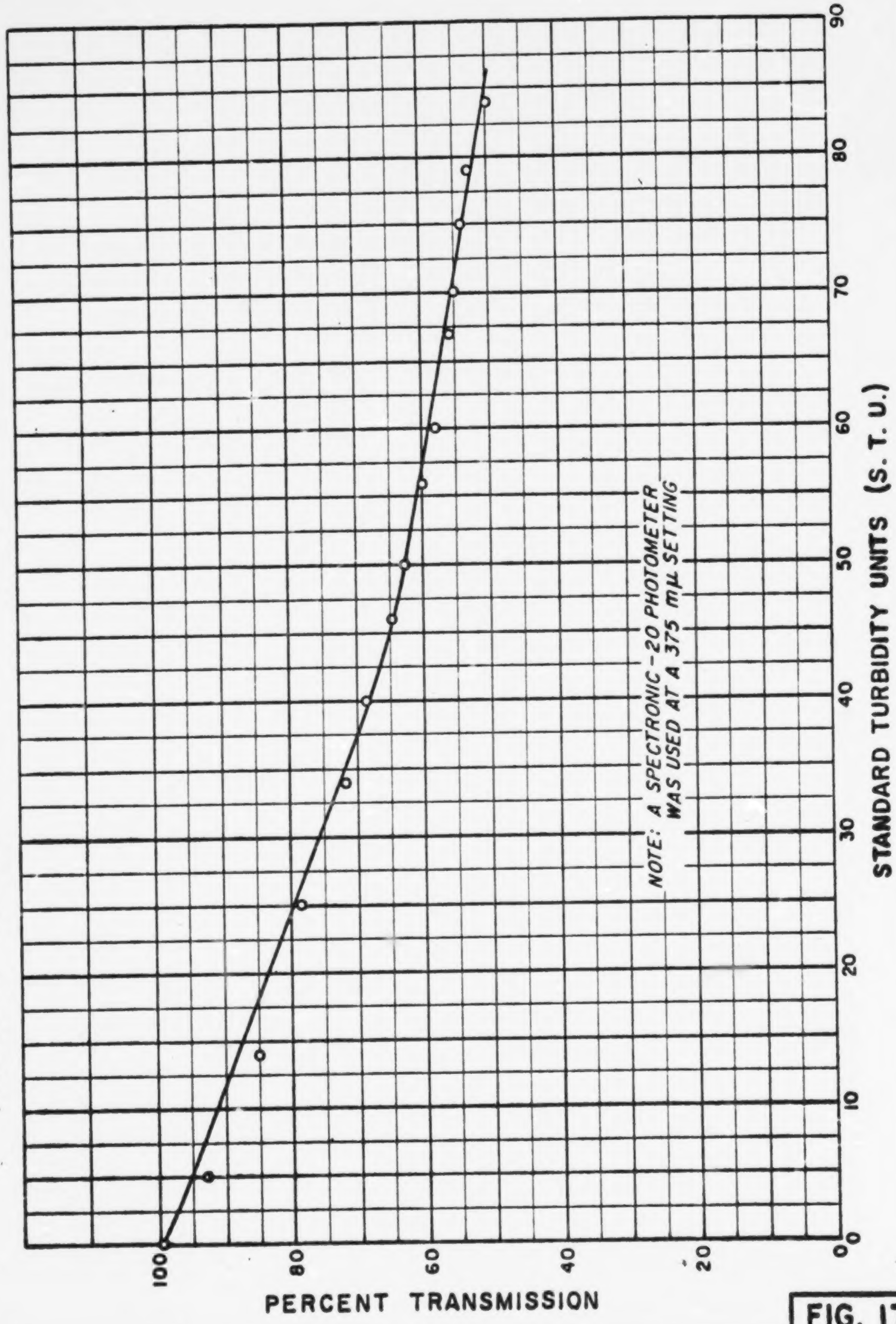


FIG. 17

TURBIDITY REDUCTION WITH HYDRATED LIME COAGULATION

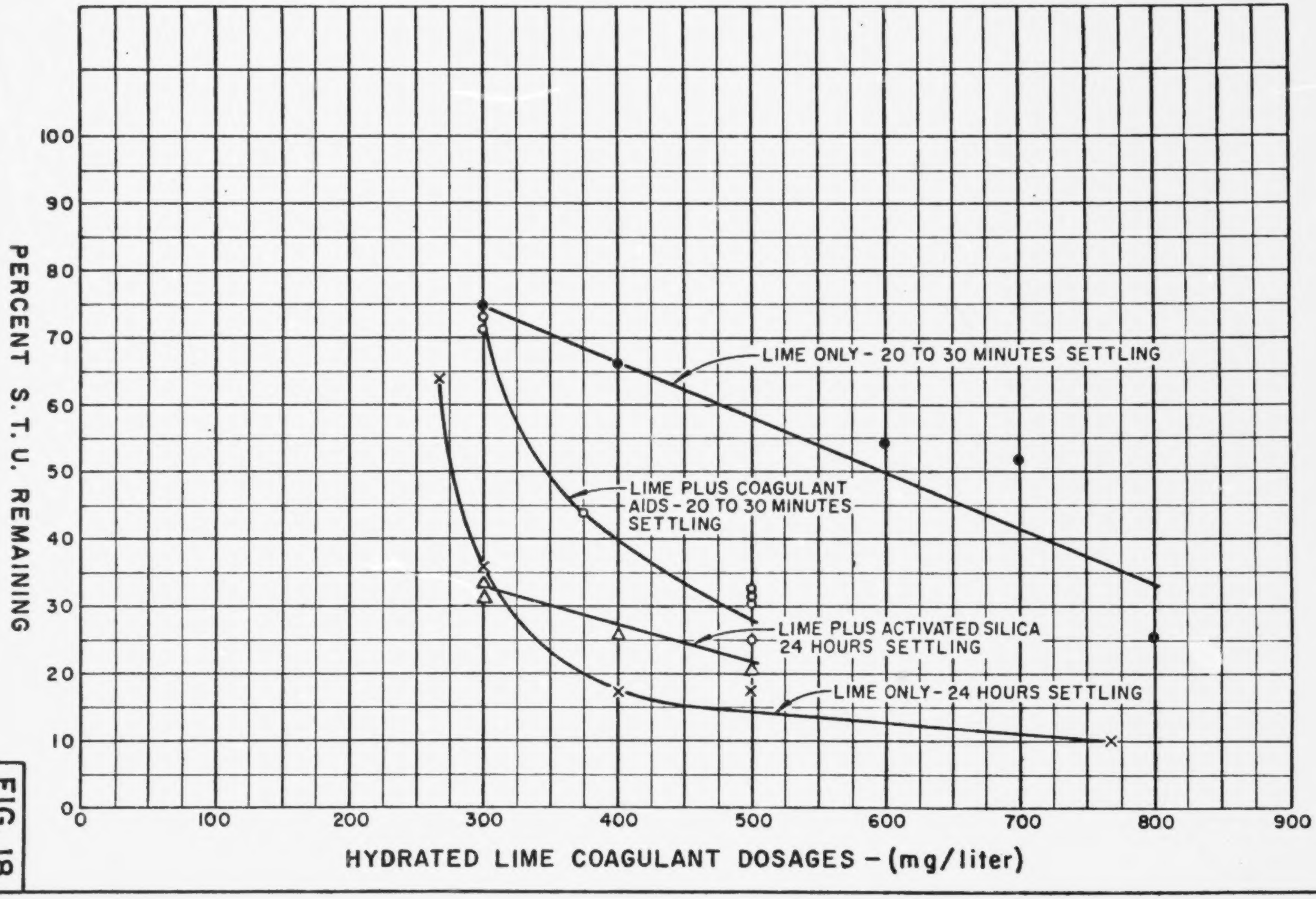


FIG. 18

TURBIDITY REDUCTION WITH ALUMINUM SULFATE COAGULATION

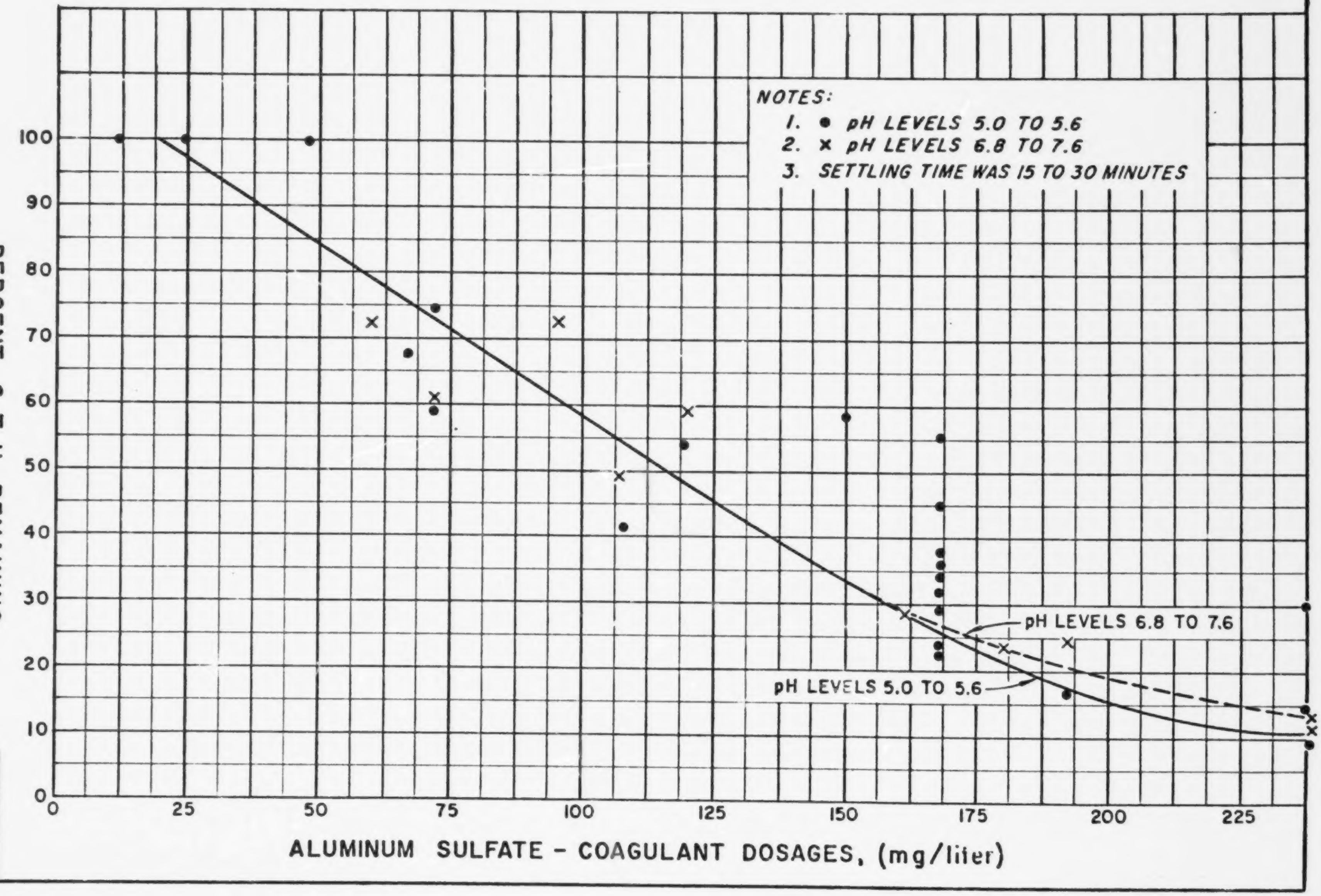


FIG. 19

TURBIDITY REDUCTION WITH FERRIC SULFATE COAGULATION

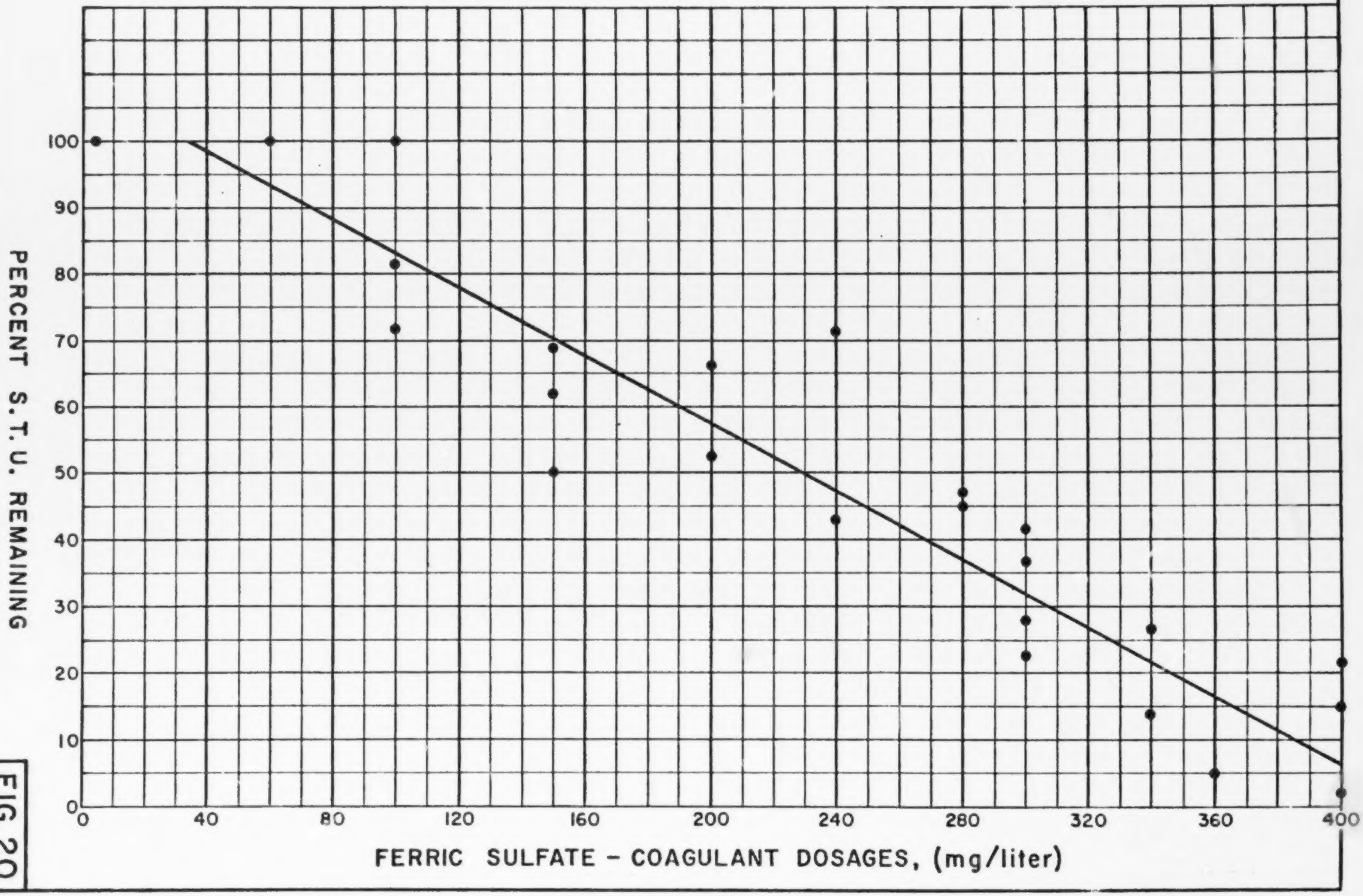
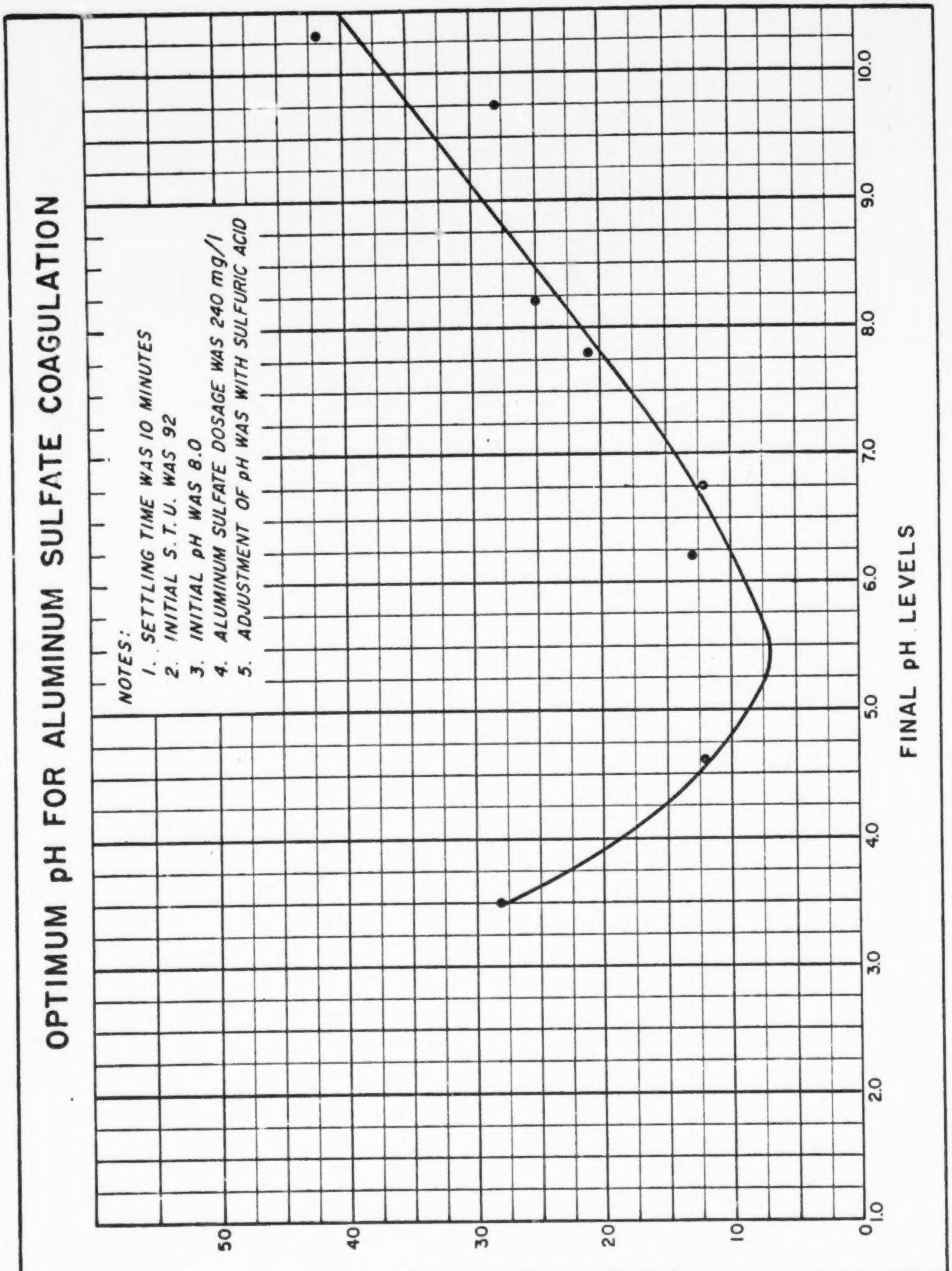


FIG. 20



OPTIMUM pH FOR ALUMINUM SULFATE COAGULATION

NOTES:
1. SETTLING TIME WAS 10 MINUTES
2. INITIAL S.T.U. WAS 92
3. INITIAL pH WAS 8.0
4. ALUMINUM SULFATE DOSAGE WAS 240 mg/l
5. ADJUSTMENT OF pH WAS WITH SULFURIC ACID

FIG. 21

EFFECT OF INITIAL ALGAE TURBIDITY ON ALUMINUM SULFATE COAGULATION

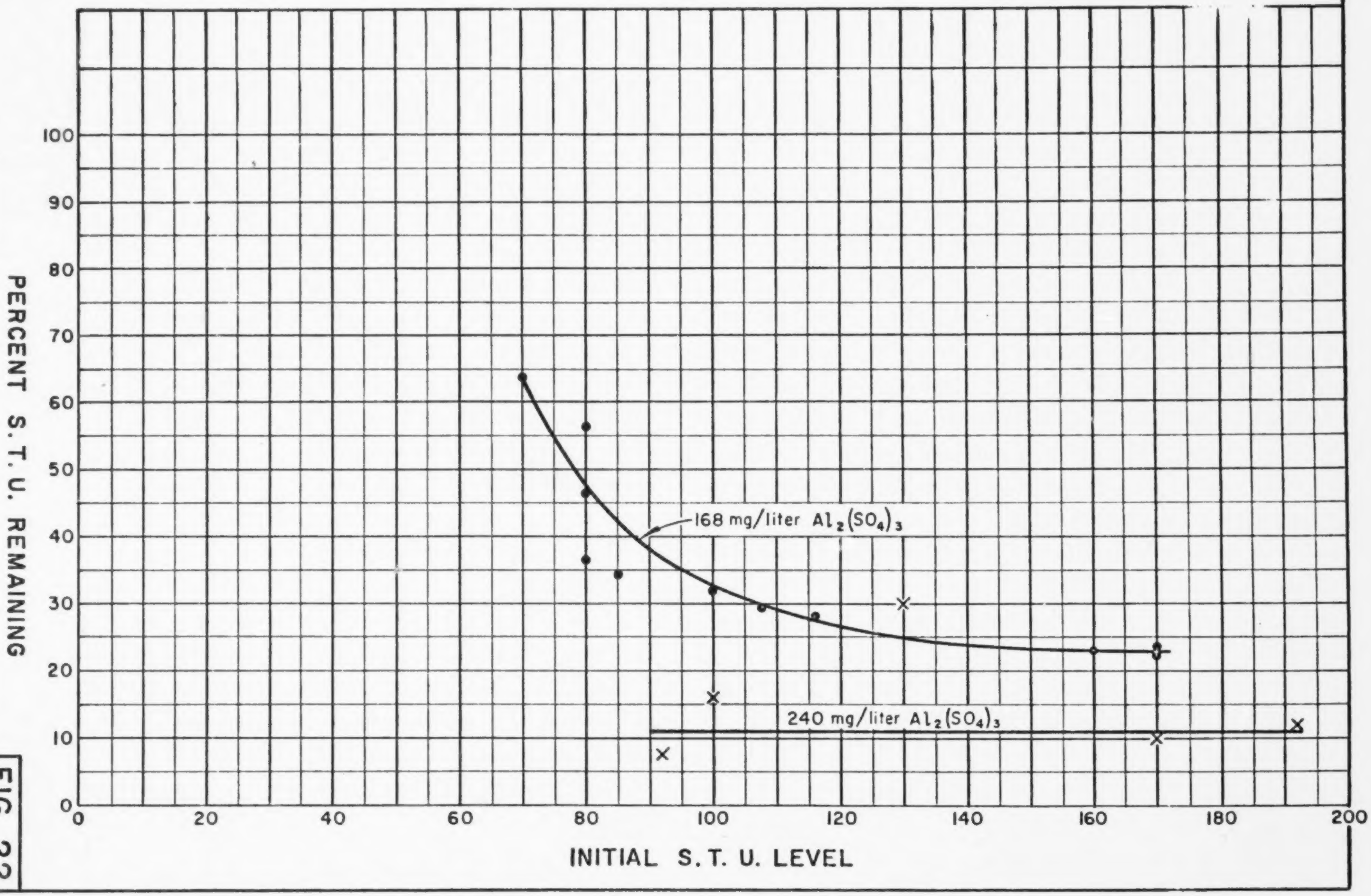


FIG. 22

EFFECT OF INITIAL ALGAE TURBIDITY ON ALUMINUM SULFATE COAGULATION

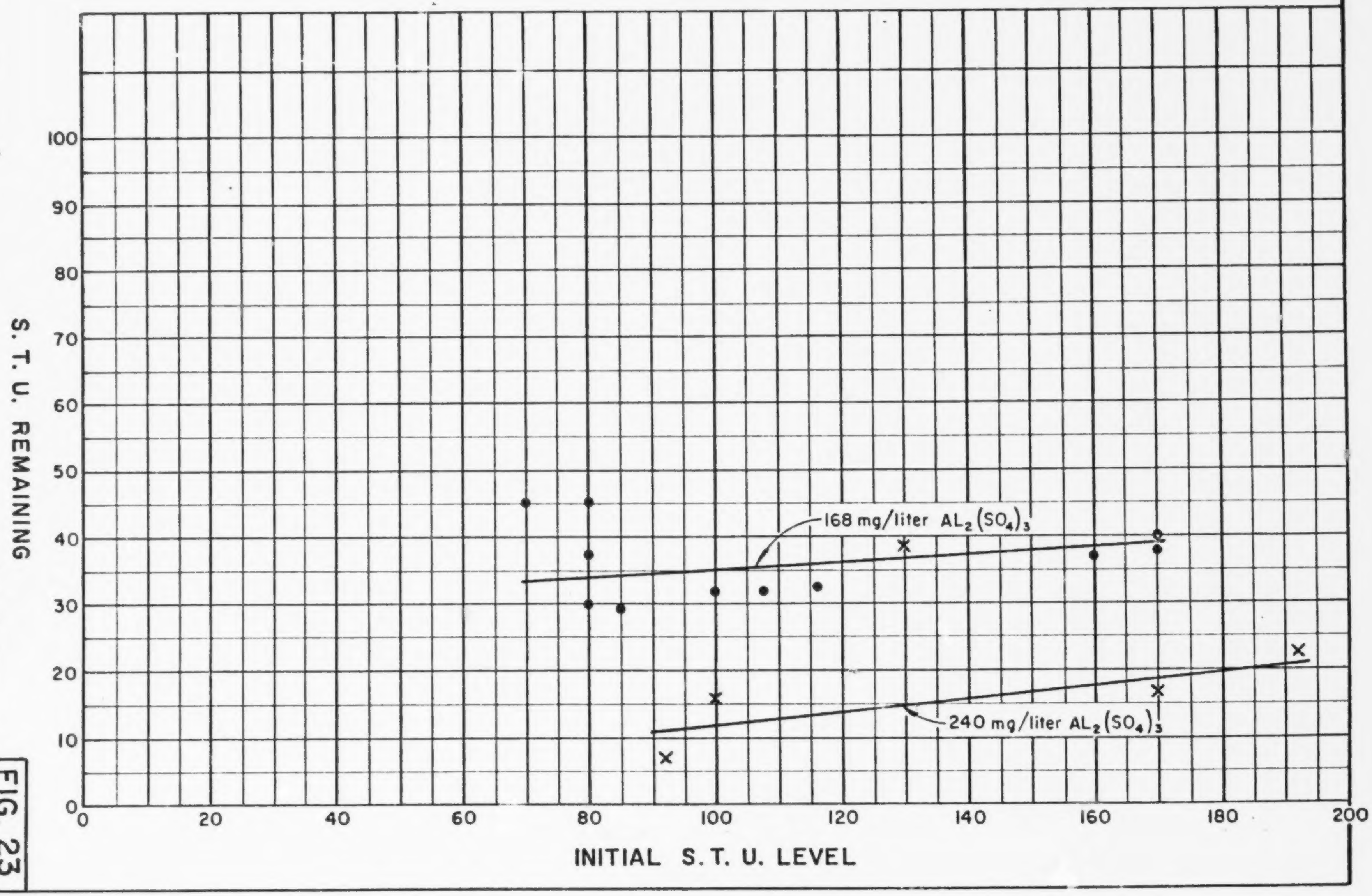


FIG. 23

TURBIDITY REDUCTION WITH ALUMINUM SULFATE COAGULATION

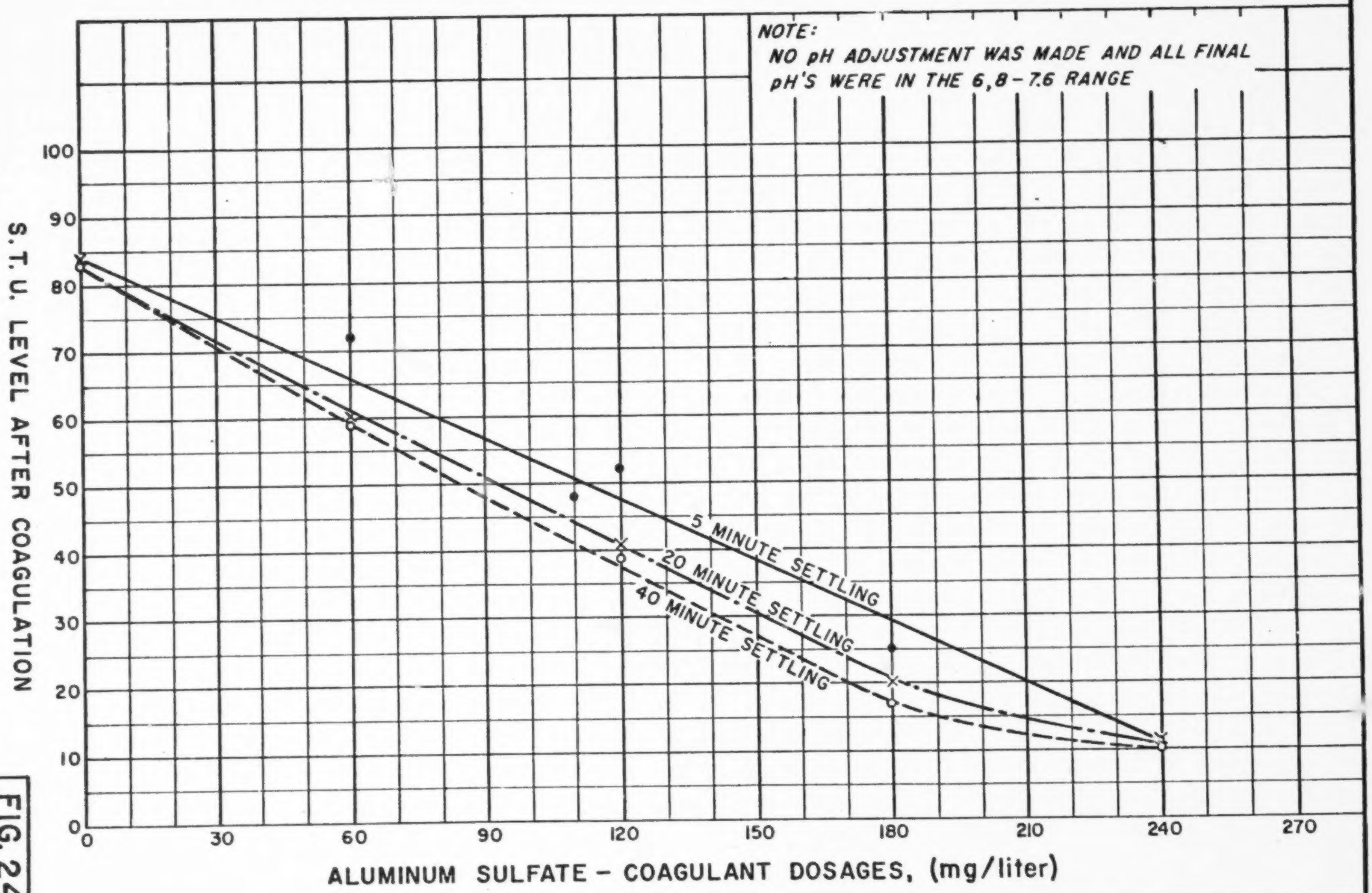


FIG. 24

TURBIDITY REDUCTION WITH ALUMINUM SULFATE COAGULATION

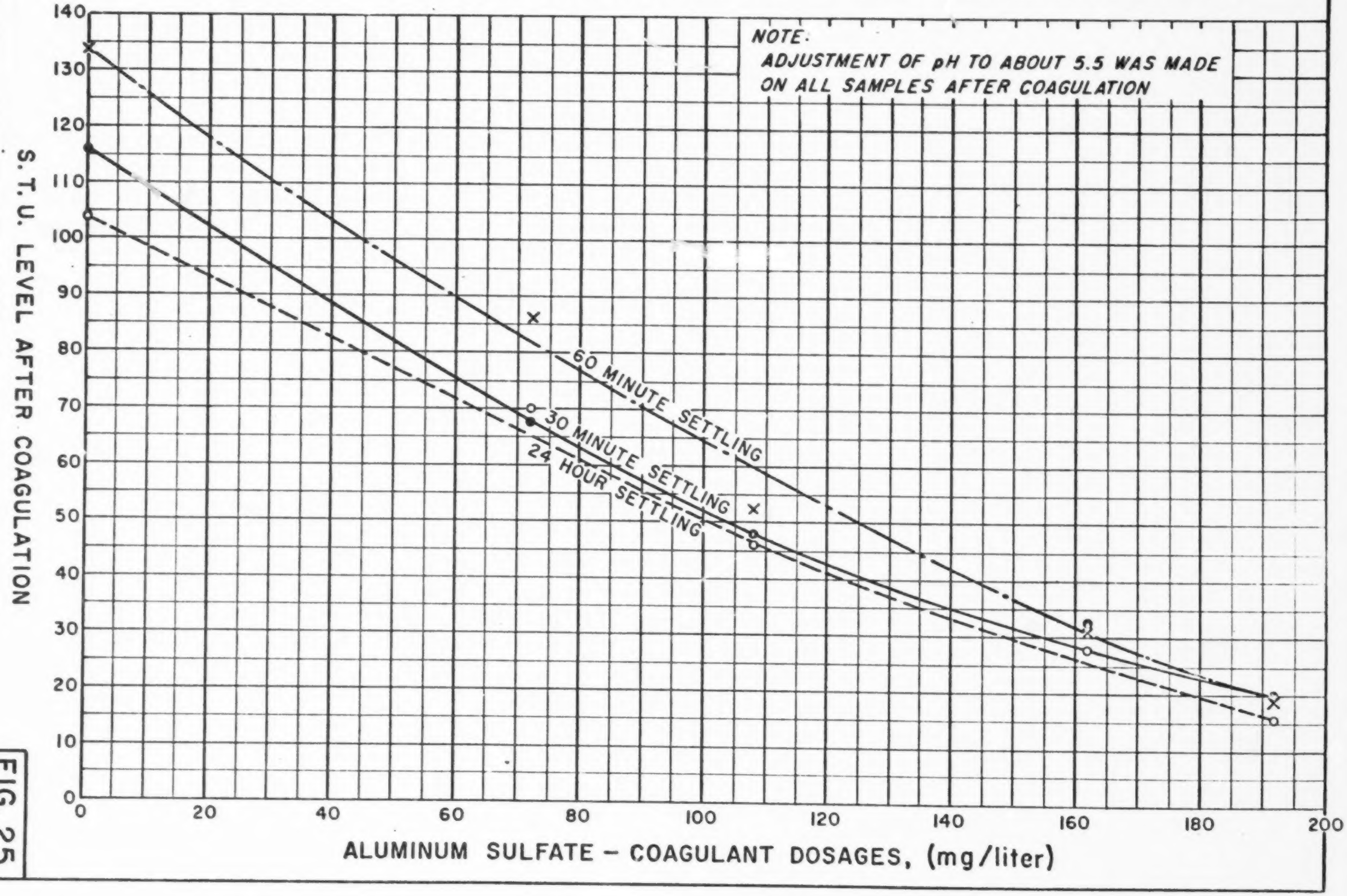


FIG. 25

POND 3 - PHOSPHATE REDUCTION BY CALCIUM HYDROXIDE COAGULATION

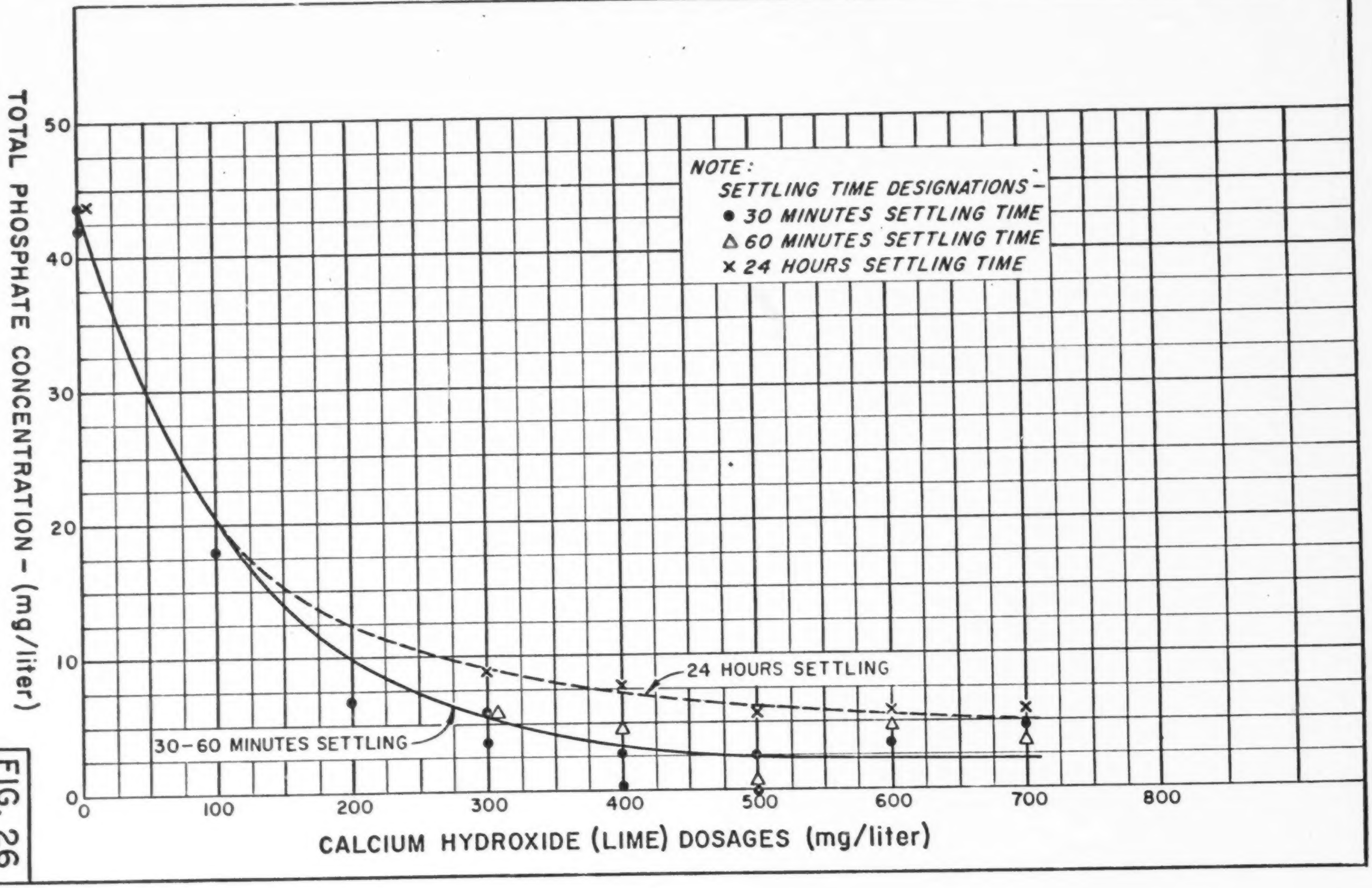


FIG. 26

POND 3 - PHOSPHATE REDUCTION WITH ALUMINUM SULFATE COAGULATION

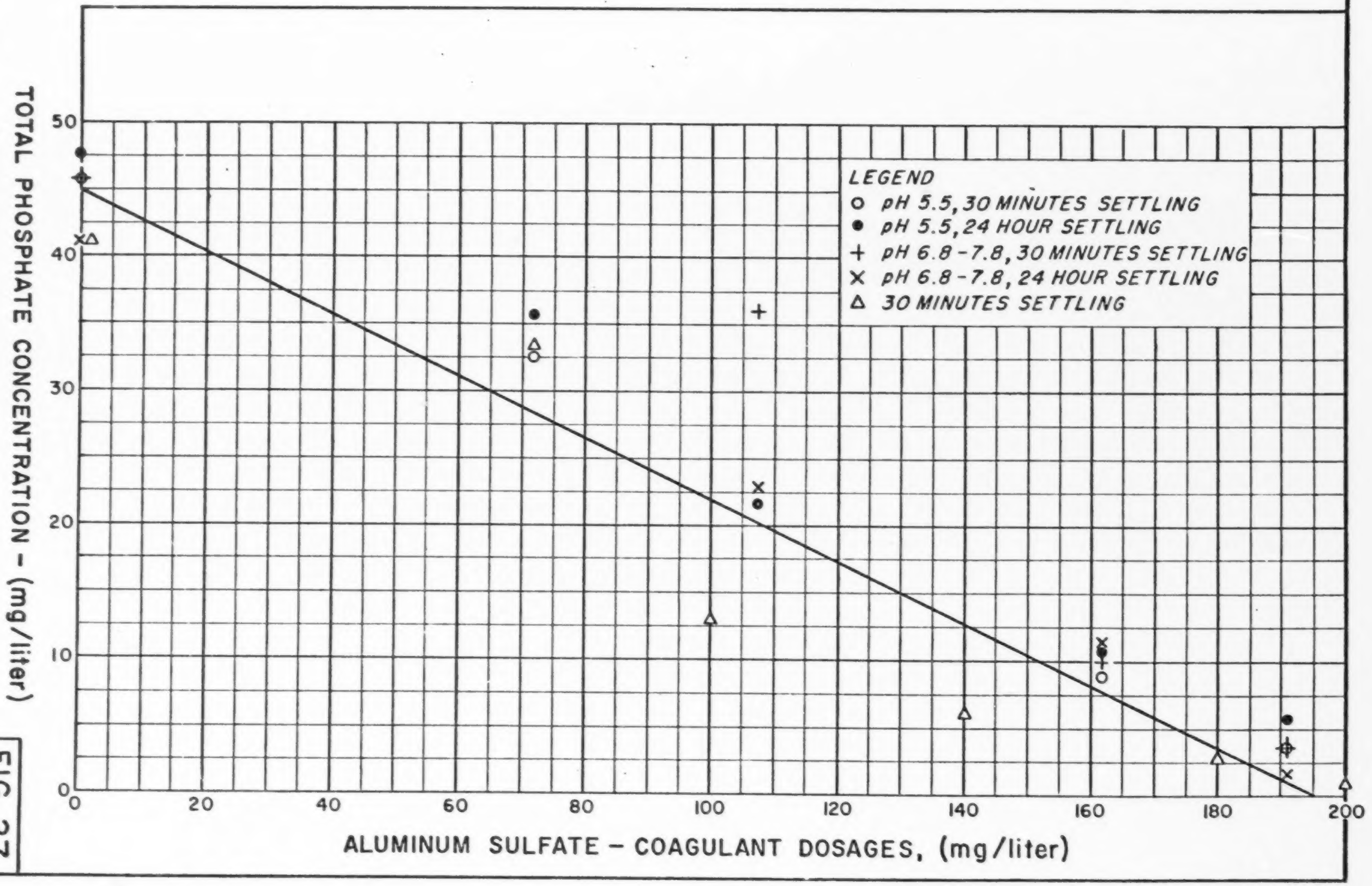


FIG. 27

POND 3 - PHOSPHATE REDUCTION BY FERRIC SULFATE COAGULATION

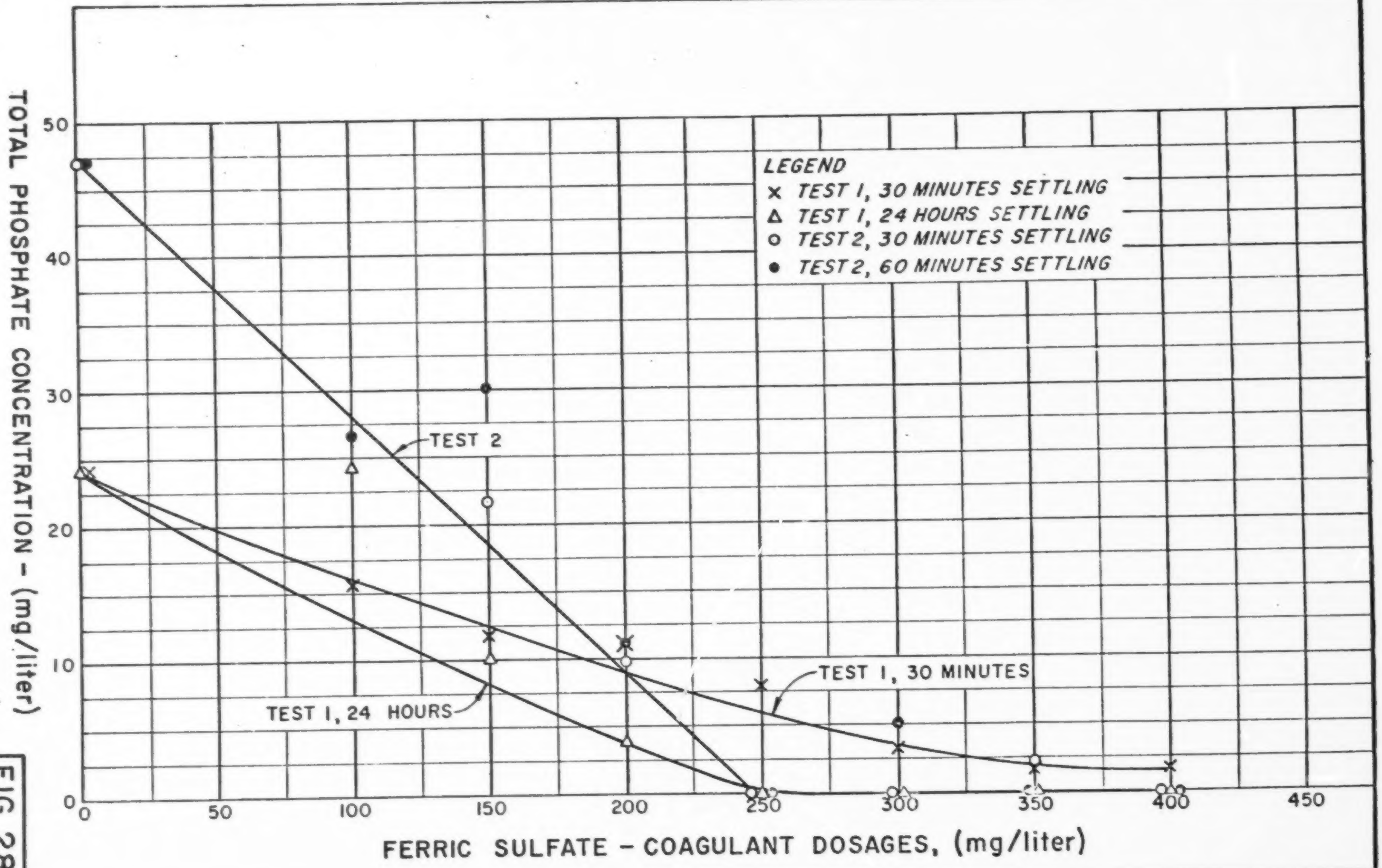


FIG. 28

COST VS. REMOVAL EFFICIENCY FOR PRIMARY COAGULANTS
BASED ON LABORATORY JAR TESTS

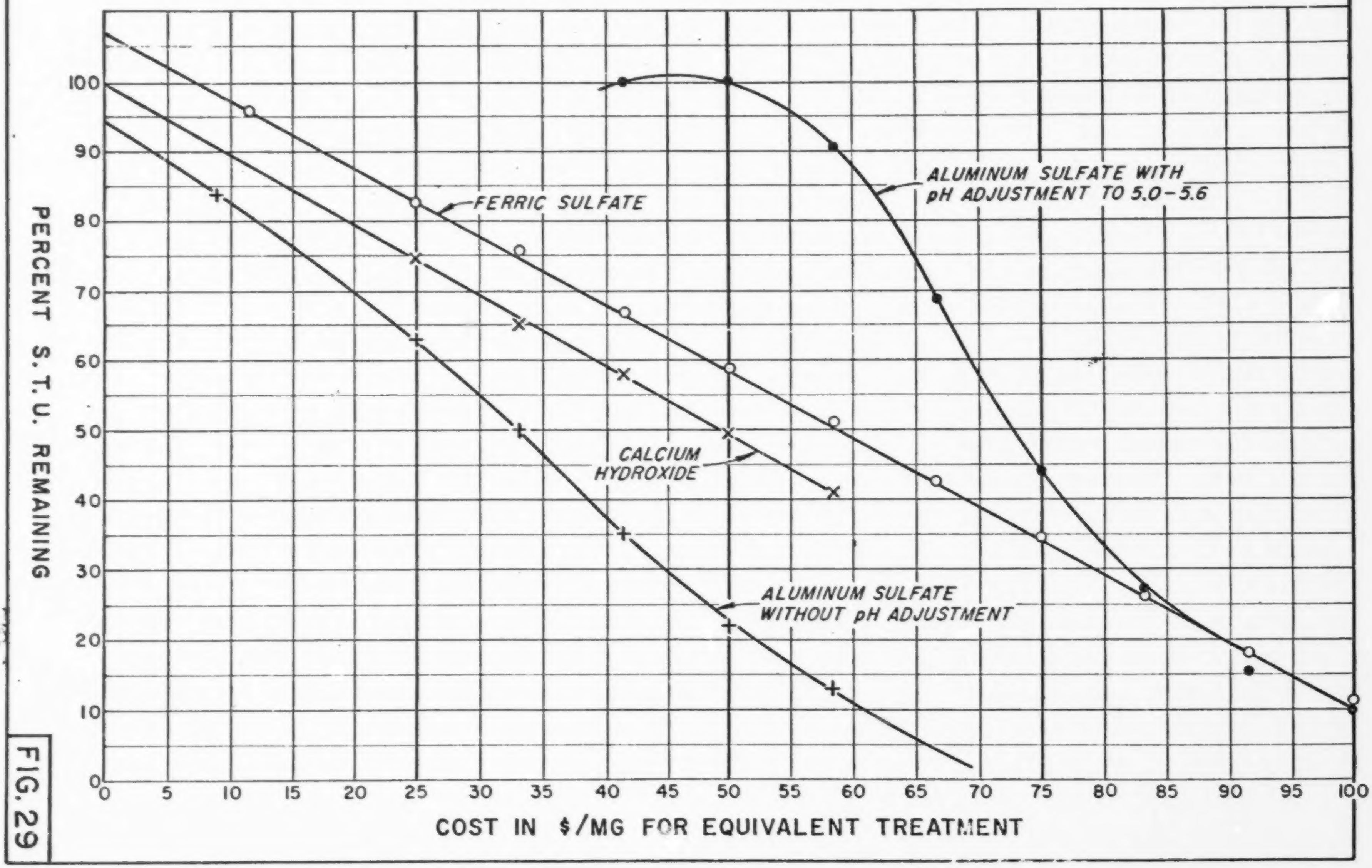
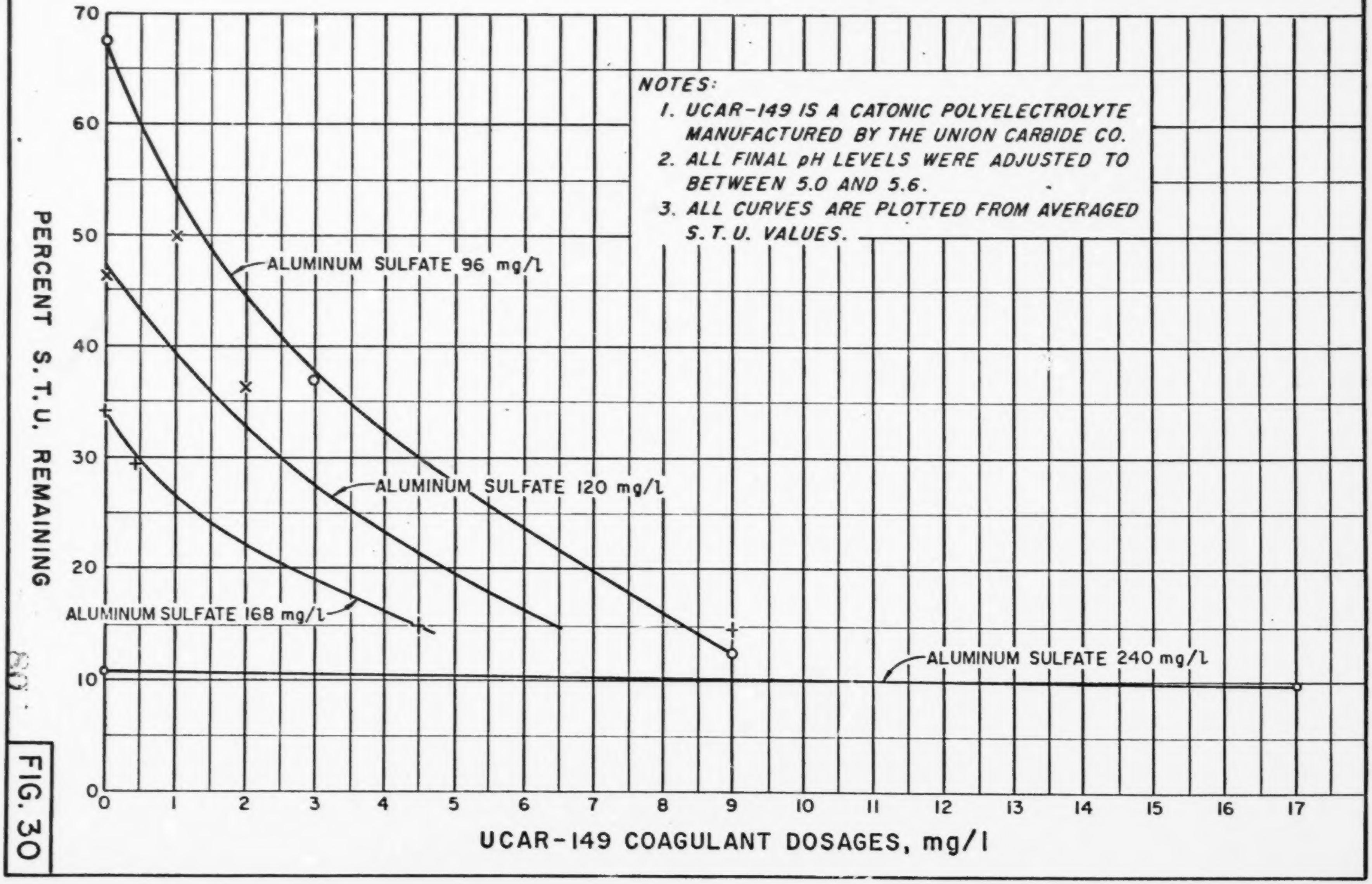


FIG. 29

TURBIDITY REDUCTION WITH ALUMINUM SULFATE AND UCAR-149



NOTES:

1. *UCAR-149 IS A CATONIC POLYELECTROLYTE MANUFACTURED BY THE UNION CARBIDE CO.*
2. *ALL FINAL pH LEVELS WERE ADJUSTED TO BETWEEN 5.0 AND 5.6.*
3. *ALL CURVES ARE PLOTTED FROM AVERAGED S.T.U. VALUES.*

FIG. 30

DENSOPAC UNIT
TURBIDITY REDUCTION WITH HYDRATED LIME COAGULATION

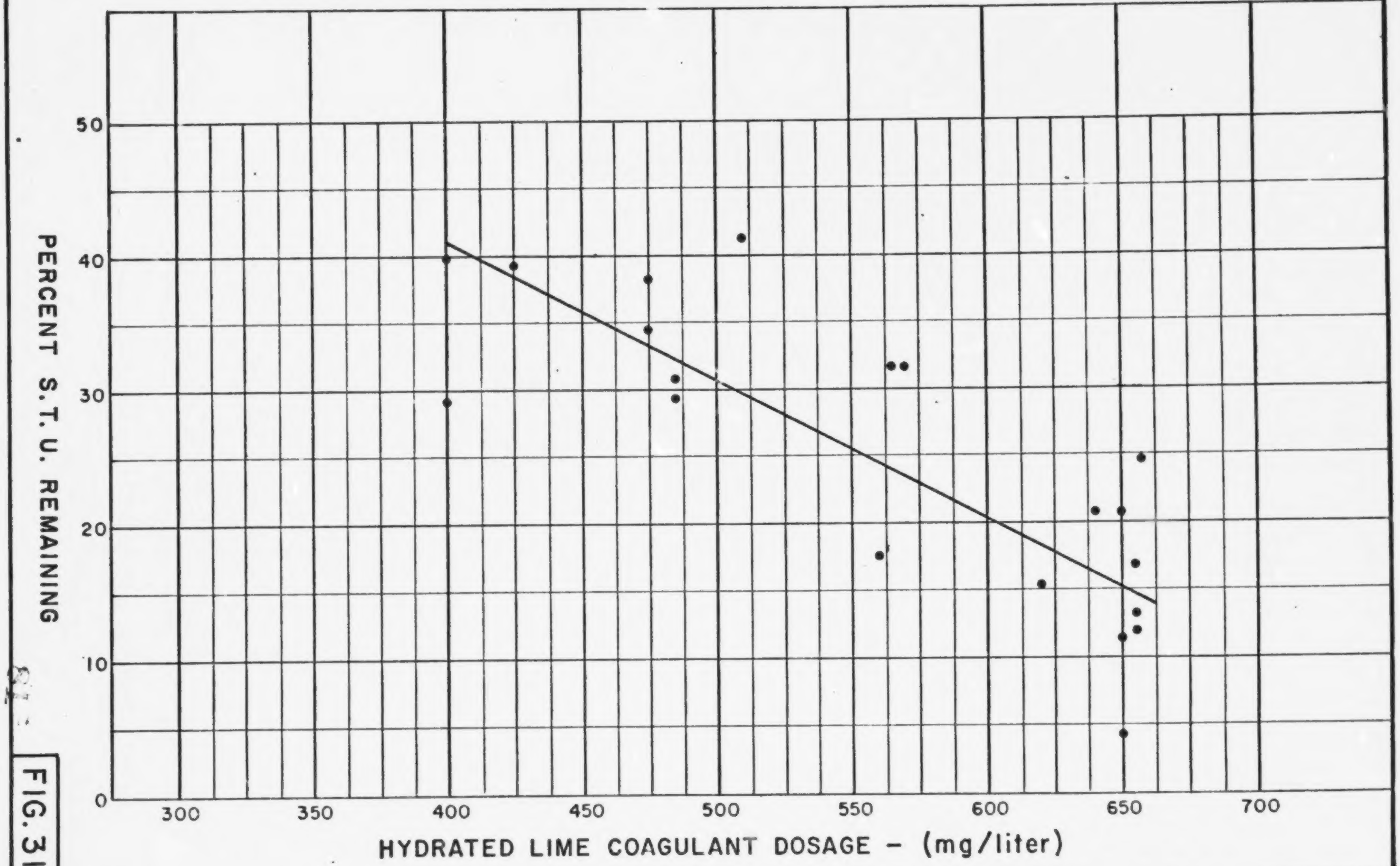


FIG. 31

FLOTATION UNIT

TURBIDITY REDUCTION WITH HYDRATED LIME COAGULATION

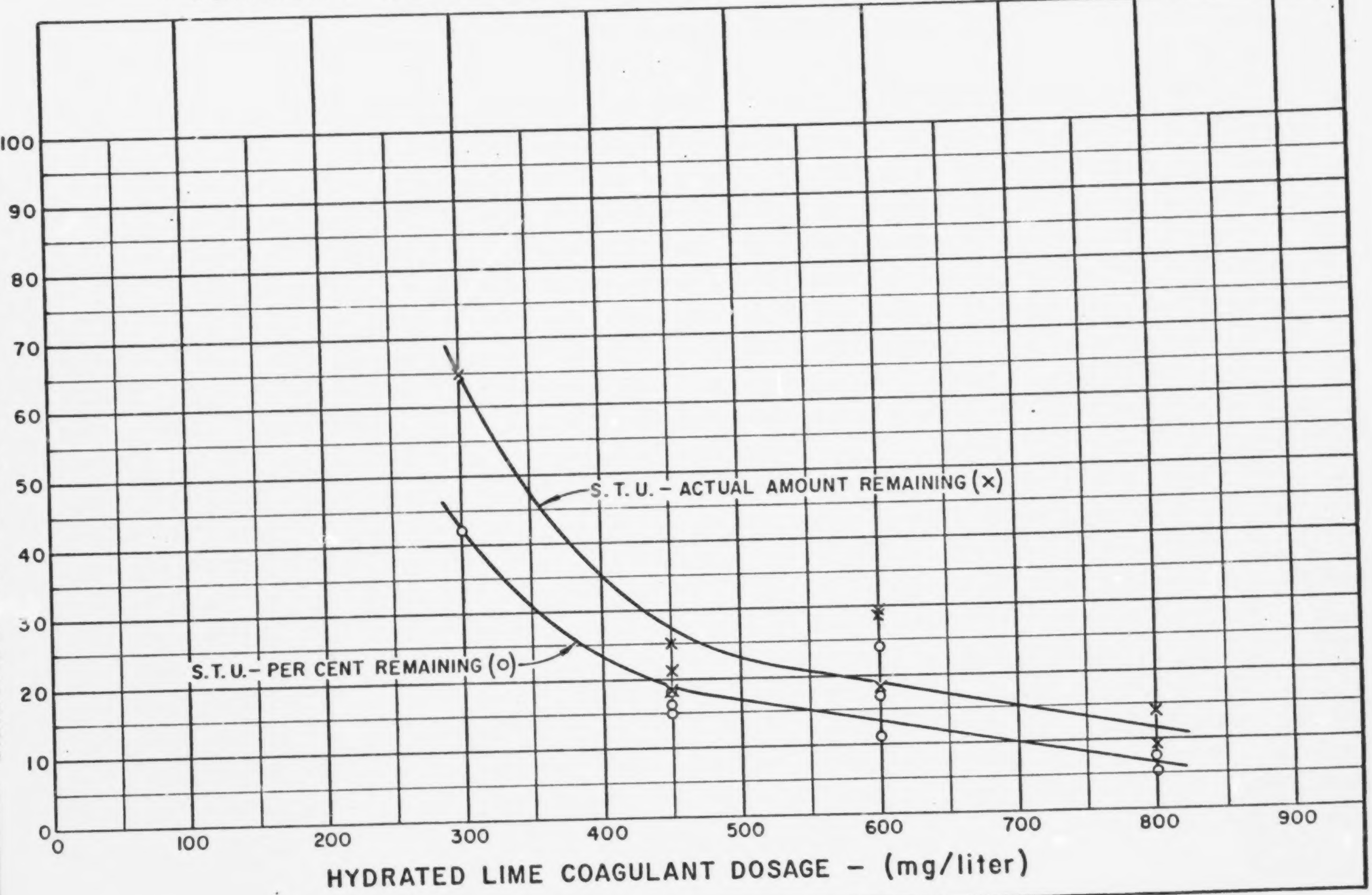


FIG. 32

--OCTOBER--

Sample	AMMONIA NITROGEN			TOTAL PHOSPHATE			C.O.D.			TOTAL SOLIDS			SUS. SOLIDS			DIS. SOLIDS		
	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.
Pr. Eff.	27.8	35.0	24.4	24	25	22	242	288	202	449	498	437	78	99	54	371	393	350
Pond 1	8.8	10.4	6.0	38	49	30	301	469	216	713	909	630	91	103	75	622	615	542
Pond 2	7.5	8.9	6.0	38	42	32	253	289	226	649	693	619	81	118	60	568	578	539
Pond 4	6.1	7.5	4.0	33	34	30	226	272	198	656	760	625	79	121	57	577	638	545
Pond 3	3.0	4.9	1.0	33	35	30	167	184	123	630	657	611	56	69	45	574	607	542
F. Eff.	2.8	3.9	1.9	36	38	28	142	186	125	664	708	642	49	49	34	615	644	595

--NOVEMBER--

	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.
Pr. Eff.	28.9	33.2	24.4	23	27	19	313	353	264	547	575	498	100	112	87	447	463	411
Pond 1	10.7	11.6	9.1	37	41	33	271	314	236	641	664	596	106	149	79	535	585	472
Pond 2	10.4	12.0	9.0	35	37	34	232	236	225	624	646	596	90	113	74	534	562	492
Pond 4	10.0	10.9	10.0	36	42	31	220	236	194	631	637	593	92	106	85	539	552	508
Pond 3	6.7	7.4	5.3	38	48	31	210	233	186	603	623	566	73	112	42	530	566	511
F. Eff.	2.5	3.0	2.3	34	42	28	175	209	149	603	658	576	75	117	46	528	595	506

--DECEMBER--

	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.
Pr. Eff.	25.2	27.3	21.2	25	28	19	343	397	285	534	636	481	125	174	95	409	499	307
Pond 1	14.6	16.1	14.0	39	41	35	254	270	236	625	638	567	98	107	91	527	540	460
Pond 2	12.6	14.7	10.8	34	37	30	224	270	193	587	607	565	89	114	73	498	526	464
Pond 4	12.2	13.5	11.2	36	40	30	240	320	199	605	678	531	100	130	80	505	548	451
Pond 3	9.2	10.2	8.8	35	38	30	209	275	178	591	620	556	69	80	58	522	543	498
F. Eff.	4.1	4.9	3.6	32	37	24	182	193	174	625	646	601	49	57	43	576	603	555

TABLE 1: CHEMICAL ANALYSES OF NUTRIENTS AND OTHER CONSTITUENTS FROM OCTOBER THROUGH DECEMBER, 1963

--JANUARY--

Sample	AMMONIA NITROGEN Av.	Max.	Min.	TOTAL PHOSPHATE Av.	Max.	Min.	C.O.D. Av.	Max.	Min.	TOTAL SOLIDS Av.	Max.	Min.	SUS. SOLIDS Av.	Max.	Min.	DIS. SOLIDS Av.	Max.	Min.
Pr. Eff.	24.9	26.2	22.4	31	34	27	369	450	297	469	594	383	144	220	76	325	374	289
Pond 1	16.3	19.6	14.0	43	45	41	285	297	267	623	650	584	114	156	32	509	516	442
Pond 2	16.3	19.6	13.2	41	42	40	210	221	199	582	608	538	82	102	52	500	529	480
Pond 4	14.9	18.0	13.2	41	42	39	208	213	199	560	598	528	80	90	72	480	514	438
Pond 3	11.7	13.2	10.8	40	42	37	199	217	184	579	624	520	56	70	40	523	584	466
F. Eff.	6.0	6.0	6,0	38	41	36	187	-	-	582	622	520	75	122	42	507	563	398

--FEBRUARY--

Sample	AMMONIA NITROGEN Av.	Max.	Min.	TOTAL PHOSPHATE Av.	Max.	Min.	C.O.D. Av.	Max.	Min.	TOTAL SOLIDS Av.	Max.	Min.	SUS. SOLIDS Av.	Max.	Min.	DIS. SOLIDS Av.	Max.	Min.
Pr. Eff.	32.6	38.5	29.0	36	49	28	408	560	312	569	642	459	66	84	53	503	580	406
Pond 1	17.6	18.0	16.8	42	43	41	293	342	268	576	675	474	83	144	50	493	531	424
Pond 2	17.6	20.5	17.4	36	40	33	307	339	290	630	640	609	76	118	48	554	579	522
Pond 4	14.8	16.0	14.0	41	45	39	266	274	262	603	620	583	79	152	32	524	553	468
Pond 3	14.0	15.5	14.0	39	43	33	222	236	202	567	614	483	57	106	27	510	566	508
F. Eff.	3.8	4.8	3.2	37	40	35	191	203	179	626	632	622	49	108	17	577	615	514

--MARCH--

Sample	AMMONIA NITROGEN Av.	Max.	Min.	TOTAL PHOSPHATE Av.	Max.	Min.	C.O.D. Av.	Max.	Min.	TOTAL SOLIDS Av.	Max.	Min.	SUS. SOLIDS Av.	Max.	Min.	DIS. SOLIDS Av.	Max.	Min.
Pr. Eff.	20.5	-	-	24	26	22	295	308	282	462	463	461	48	50	45	414	413	406
Pond 1	15.0	-	-	44	45	43	352	386	317	645	670	620	77	83	70	570	587	568
Pond 2	15.0	-	-	42	44	39	286	289	282	617	632	602	64	64	64	553	568	538
Pond 4	11.5	-	-	47	48	45	370	457	282	699	748	650	89	120	57	610	628	593
Pond 3	12.8	-	-	47	48	46	404	529	279	717	770	664	95	143	46	622	627	618
F. Eff.	1.2	-	-	39	43	34	175	188	162	654	-	-	17	-	-	637	-	-

TABLE 2: CHEMICAL ANALYSES OF NUTRIENTS AND OTHER CONSTITUENTS FROM JANUARY THROUGH MARCH, 1964

—APRIL—

Sample	AMMONIA NITROGEN			TOTAL PHOSPHATE			C.O.D.			TOTAL SOLIDS			SUS. SOLIDS			DIS. SOLIDS		
	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.	Av.	Max.	Min.
Pr. Eff.	30.0	-	-	60	-	-	588	-	-	676	-	-	98	-	-	583	-	-
Pond 1	16.8	-	-	52	-	-	570	-	-	797	-	-	175	-	-	622	-	-
Pond 2	18.0	-	-	42	-	-	236	-	-	597	-	-	65	-	-	532	-	-
Pond 4	17.0	-	-	48	-	-	440	-	-	758	-	-	139	-	-	619	-	-
Pond 3	17.4	-	-	48	-	-	440	-	-	756	-	-	114	-	-	642	-	-
F. Eff.	2.4	-	-	36	-	-	158	-	-	633	-	-	26	-	-	607	-	-

TABLE 3: CHEMICAL ANALYSES OF NUTRIENTS AND OTHER CONSTITUENTS FOR APRIL, 1964

TABLE 4

STATISTICAL DISTRIBUTION ANALYSES OF POND 3

Sample No.	Total Solids (mg/l)	Suspended Solids (mg/l)	Dissolved Solids (mg/l)	Ammonia Nitrogen (mg/l)	Total Phosphate (mg/l)
1S	644	82	562	14.0	38
1L	616	22	594	13.6	37
2M	604	48	556	13.3	38
2L	623	58	565	13.6	37
3S	641	87	554	12.8	37
3M	594	44	550	12.0	38
4S	641	134	507	11.6	38
4L	600	36	564	12.0	38
5S	613	56	557	12.8	37
5M	597	34	563	12.4	40
5L	608	49	559	11.6	38
6M	586	34	552	11.2	40
6L	632	46	586	13.2	38
7S	655	146	509	13.6	41
7L	640	22	618	12.8	39
8S	704	162	542	12.4	40
8M	598	54	544	12.0	39
9M	596	64	532	12.0	38
Inf.	671	66	605	12.8	41
Eff.	613	94	519	11.2	36

Notes: Samples were taken from 11:00 AM to 12:30 PM on December 31, 1963.

TABLE 5

STATISTICAL DISTRIBUTION ANALYSES OF POND 3

Sample No.	Total Solids (mg/l)	Suspended Solids (mg/l)	Dissolved Solids (mg/l)	Ammonia Nitrogen (mg/l)	Total Phosphate (mg/l)	Dissolved Oxygen (mg/l)
1S	826	36	790	10.0	28	4.8
1L	772	62	710	9.8	31	4.4
2M	590	42	548	9.8	29	4.1
2L	628	56	572	7.8	30	4.1
3S	604	50	554	9.4	32	4.2
3M	606	70	536	8.2	30	4.7
4S	734	50	684	11.6	31	4.4
4L	702	98	604	11.2	32	4.1
5S	660	34	626	9.0	32	5.0
5M	736	72	664	9.8	32	4.6
5L	624	32	592	9.8	30	5.6
6M	616	54	562	9.8	30	3.8
6L	750	78	672	10.0	21	4.6
7S	632	66	566	9.8	21	4.2
7L	636	48	588	9.8	22	4.2
8S	698	70	628	10.4	27	3.3
8M	692	72	620	7.8	29	3.8
9M	616	80	536	8.6	30	4.4
Inf.	758	94	664	12.0	27	3.4
Eff.	648	54	594	7.8	31	6.3

Notes: Samples were taken from 10:00 to 11:30 AM on January 7, 1964.

TABLE 6

STATISTICAL DISTRIBUTION ANALYSES OF POND 3

Sample No.	Total Solids (mg/l)	Suspended Solids (mg/l)	Dissolved Solids (mg/l)	Ammonia Nitrogen (mg/l)	Total Phosphate (mg/l)	Dissolved Oxygen (mg/l)
1S	607	62	545	12.4	42	7.5
1L	579	42	537	12.8	40	9.1
2M	589	80	509	12.0	39	7.0
2L	609	32	577	12.0	43	8.8
3S	646	90	556	11.2	41	9.1
3M	644	64	580	12.4	37	8.5
4S	570	56	514	10.8	39	9.1
4L	598	28	570	11.2	39	9.6
5S	583	72	511	10.0	41	7.0
5M	597	38	559	10.4	40	7.2
5L	606	22	584	9.8	42	7.2
6M	578	36	542	10.0	43	8.5
6L	611	26	585	10.8	43	7.0
7S	601	-	-	-	40	6.8
7L	627	-	-	-	39	6.4
8S	626	-	-	-	41	6.1
8M	592	-	-	-	41	5.9
9M	611	-	-	-	40	6.5
Inf.	585	-	-	-	40	2.1
Eff.	593	-	-	-	40	5.8

Notes: Samples 1S through 6L were taken from 3:30 to 4:30 PM on January 16, 1964. Samples 7S through Eff. were taken from 4:00 to 5:00 PM on January 17, 1964.

TABLE 7

SUMMARY OF CHEMICAL COSTS

Chemical	Unit Cost	How Packaged	Cost per Pound	Cost per Million Gallons 1 mg/1	10 mg/1	100 mg/1	Source of Cost Estimate
High-Ca++ Quicklime	$29.15/ton 33.40/ton	Bulk 50 lb. bags	$0.0146 0.0167	$ 0.122 0.139	$ 1.22 1.39	$ 12.20 13.90	U.S. Lime Products
High-Ca++ Hyd. Lime	27.80/ton 30.50/ton	Bulk 50 lb. bags	0.0139 0.0153	0.116 0.127	1.16 1.27	11.60 12.70	U.S. Lime Products U.S. Lime Products
Local Hyd. Lime	23.40/ton 20.00/ton	50 lb. bags Bulk	0.012 0.010	0.10 0.0834	1.00 0.834	10.00 8.34	Stauffer Chemical (
Sulfuric Acid	0.03/lb. 0.0175/lb.	55 gal. drum Tank truck	0.030 0.175	0.250 0.146	2.50 1.46	25.00 14.60	McKesson & Robbins McKesson & Robbins
Aluminum Sulfate	3.75/100 lbs. 3.00/100 lbs.	bags-1 tons bags-10 tons	0.0375 0.030	0.313 0.250	3.13 2.50	31.30 25.00	McKesson & Robbins McKesson & Robbins
$FeSO_4$	55.00/ton	bags-5000 lbs.	0.0275	0.229	2.29	22.90	McKesson & Robbins
$CuSO_4$	14.10/100 lb. 13.80/100 lb.	5 ton lots 10 ton lots	0.141 0.138	1.172 1.150	11.72 11.50	117.20 115.00	McKesson & Robbins McKesson & Robbins
$FeCl_3$	5.00/100 lb.	10-55 gal. drums	0.05	0.417	4.17	41.70	McKesson & Robbins
$Na_2Al_2O_3$	0.12/lb.	100 lb. bags	0.12	1.00	10.00	100.00	McKesson & Robbins
$CaCl_2$	68.20/ton	100 lb. bags	0.034	0.283	2.83	28.30	McKesson & Robbins
UCAR-149	1.00-1.25/lb.	25-50 lb. pkg.	1.00	8.34	83.40	834.00	Union Carbide
Superfloc-16	1.25/lb.	25-50 lb. pkg.	1.25	10.40	104.00	1040.00	American Cyanimide
WPB-25	0.50/lb.	25-50 lb. pkg.	0.50	4.17	41.70	417.00	Stein Hall

TABLE 8

JAR TEST COAGULATION RESULTS FOR ALUMINUM SULPHATE WITH AND WITHOUT COAGULANT AIDS (1)

Run No.	Initial S.T.U.	Aluminum Sulfate mg/l	Aluminum Sulfate Only: S.T.U. After Treatment	Aluminum Sulfate Only: %S.T.U. Remaining	Aluminum Sulfate With Coagulant Aids: Coagulant Aid Type (4)	Coagulant Aid Dosage, mg/l	S.T.U. After Treatment	%S.T.U. Remaining	% Improvement Due to the Coagulant Aid	Chemical Costs per Million Gallons (5)
1	80	96	54	67.5	A	3.0	37	46.2	21.3	$ 88.50
2	72	96	52	72.0	A	3.0 to 9.0	9 (2)	12.5	59.5	88.50 to 138.50
3	80	120	37	46.3	A	1.0	40	50.0	- 3.7	77.80
4	80	120	37	46.3	A	2.0	29	36.3	10.0	86.20
5	80	120	37	46.3	A	3.0	20	25.0	21.3	94.50
6	72	120	-	-	A	6.0	7	9.6	-	119.50
7	85	168	29	34.1	A	0.1 to 0.9	25 (2)	29.4	4.7	82.30 to 89.00
8	85	168	29	34.1	A	1.0 to 9.0	15 (2)	17.7	16.4	89.90 to 156.50
9	137	168	39	28.5	B	1.0 to 11.0	34 (2)	24.8	3.7	82.00 (3
10	108	168	32	29.6	C	0.1 to 0.9	27 (2)	25.0	4.6	82.00 (3
11	80	168	37	46.3	D	0.7 to 0.9	25 (2)	31.3	15.0	82.00 (3
12	80	168	37	46.3	E	0.7 to 1.0	25 (2)	31.3	15.0	82.00 (3
13	100	168	32	32.0	F	0.1 to 0.5	30 (2)	30.0	2.0	82.00 (3
14	80	168	45	56.2	G	0.1 to 0.5	36 (2)	45.0	11.2	82.00 (3
15	112	240	12	10.7	A	1.0 to 17.0	11 (2)	9.8	0.9	107.90 to 241.50
16	170	240	17	10.0	B	0.1 to 11.0	20	11.8	- 1.8	100.00

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(1) All tests were performed on water from oxidation pond 3 after pH adjustment with sulfuric acid to between 5.0 and 5.6. Samples for each run were divided after receiving aluminum sulfate dosage and half the original portion received a dose of coagulant aid. Settling times varied from 20 to 40 minutes.

(2) Average values.

(3) Exclusive of coagulant aids

(4) Coagulant Aid Type Identification:

Symbol	Name	Class	Manufacturer
A	UCAR-149	cationic polyelectrolyte	Union Carbide Co.
B	Harmaco-10	pregelatinized starch	Morningside Paisley Co.
C	Perramid	nonionic polymer	Morningside Paisley Co.
D	MRL-19	cationic polyelectrolyte	Stein Hall Co.
E	MRL-22	cationic polyelectrolyte	Stein Hall Co.
F	Reten-205 MX	cationic polyelectrolyte	Morningside Paisley Co.
G	Reagent-52	cationic polyelectrolyte	Stein Hall Co.

Other coagulant aids used without significant effect were: Stein Hall products: MRL-113 and MRL-115 (anionic), WPB (nonionic), MRL-51 (cationic), MRL-13 and MRL-14 (cationic); activated sludge and activated silica.

(5) Chemical costs include $39.50 per million gallons for 270 mg/l of sulfuric acid for pH adjustment.

TABLE 9

SUMMARY - DENSOPAC UNIT TEST RESULTS

Test Run No.	Influent			Densator						Recarbonator					Total Chemical Cost Per Million Gallons
				Lime (1)	Coagulant Aids		Densator Effluent			Coagulant Aids (2)		Recarb. Effluent			
	gpm	pH	S.T.U.	mg/l	Type	mg/l	pH	S.T.U.	% S.T.U. Remaining	Type	mg/l	pH	S.T.U.	% S.T.U. Remaining	
1	10	8.6	170	485	$CaCl_2$ SF-16	100.0 0.5	11.9	50	29.4	-	-	11.6	68	40.0	$ 74.00
2	10	8.1	150	510	$CaCl_2$ SF-16	100.0 1.5	11.9	62	41.4	-	-	10.9	50	33.3	86.40
3	10	7.9	126	560	$CaCl_2$ SF-16	200.0 1.5	12.3	22	17.5	-	-	11.8	40	31.7	119.00
4(3)	10	8.4	104	620	SF-16	2.0	12.3	16	15.4	SF-16	1.0	11.3	23	22.2	83.00
5	10	8.1	126	650	SF-16	2.0	12.4	26	20.6	SF-16	2.0	10.4	32	25.4	96.00
6	10	8.4	158	570	SF-16	2.0	12.0	50	31.7	UCAR-149	0.5	10.1	54	34.2	72.50
7	10	8.1	126	565	SF-16	2.0	12.2	40	31.8	UCAR-149	1.0	11.5	38	30.2	76.50
8	10	8.0	150	640	SF-16	2.0	12.0	31	20.7	UCAR-149	2.0	11.4	37	24.6	91.00
9	10	7.9	126	655	SF-16	2.0	12.0	31	24.6	UCAR-149	2.0	10.4	40	31.7	92.00
10	15	8.7	176	650	SF-16	2.0	12.0	20	11.4	UCAR-149	2.0	11.4	28	15.9	91.50
11	15	8.9	170	655	SF-16	2.0	12.0	20	11.8	UCAR-149	2.0	11.6	16	9.4	92.00
12	15	8.4	120	655	SF-16	2.5	12.3	5	4.2	SF-16	2.0	12.1	18	15.0	101.50
13	15	8.2	108	655	SF-16	2.5	11.9	18	16.7	SF-16	2.0	11.4	21	19.5	101.50
14	15	8.3	153	655	WPB-25	2.0	12.0	20	13.1	SF-16	3.0	11.7	28	18.3	94.50
15	15	8.5	168	485	WPB-25	3.0	11.6	52	30.9	$Fe_2(SO_4)_3$	30.0	9.6	92	54.7	60.50
16	10	7.7	126	475	$CaCl_2$ $Fe_2(SO_4)_3$	50.0 10.0	11.7	48	38.1	-	-	9.1	41	32.5	56.50
17	10	8.1	134	475	$CaCl_2$ $Fe_2(SO_4)_3$	25.0 14.0	11.7	46	34.4	-	-	8.8	46	34.4	50.00
18	10	7.9	120	400	$CaCl_2$	100.0	11.6	35	29.2	-	-	11.1	44	36.7	61.50
19	10	7.7	104	425	$CaCl_2$	100.0	11.5	41	39.2	$Al_2(SO_4)_3$	10.0	10.4	46	42.6	66.50
20	10	7.7	126	400	$CaCl_2$	100.0	11.4	50	39.7	$Al_2(SO_4)_3$	20.0	9.5	43	34.1	66.50

NOTES:

(1) Lime dosages were estimated from alkalinity and hardness data.

(2) Coagulant aid identification: SF-16 = Superfloc 16, a non-ionic polymer of the American Cyanamide Co.; UCAR-149, a cationic polyelectrolyte of the Union Carbide Co.; and WPB-25, a non-ionic polymer of the Stein-Hall Co.

(3) From test run No. 4 on, a continuous slow wasting of sludge in the densator unit was instituted to prevent floc carryover.

TABLE 10

DENSOPAC UNIT, ALKALINITY AND HARDNESS DATA

Lime Dosage mg/1	Flow Stream	pH Range	Total Hardness As $CaCO_3$ mg/1	Alkalinities as CaCO3			
				Hydroxide mg/1	Carbonate mg/1	Bicarbonate mg/1	Sodium mg/1
400-450	Influent	7.7 - 7.9	80 - 92	0	0	284 - 288	194 - 204
	Densator	11.4 - 11.6	140	160 - 180	104 - 120	0	140 - 144
	Recarbonator	9.5 - 11.1	116 - 140	0 - 96	132 - 144	0 - 80	96 - 136
450-500	Influent	7.7 - 8.6	96 - 140	0	0	268 - 276	128 - 200
	Densator	11.6 - 11.9	216 - 236	260 - 264	100 - 152	0	156 - 176
	Recarbonator	8.8 - 11.6	136 - 228	0 - 256	120 - 192	0 - 124	148 - 186
500-550	Influent	8.6	124	0	0	264	140
	Densator	11.9	300	296	96	0	92
	Recarbonator	10.9	112	84	152	0	124
550-600	Influent	7.9 - 8.4	68 - 128	0	0	264 - 276	144 - 196
	Densator	12.0 - 12.3	216 - 424	344 - 396	88 - 104	0	24 - 212
	Recarbonator	10.1 - 11.8	60 - 276	0 - 120	104 - 232	0 - 56	0 - 216
600-655	Influent	7.9 - 8.9	72 - 120	0	0 - 56	216 - 268	148 - 200
	Densator	12.0 - 12.4	284 - 408	464 - 500	72 - 164	0	188 - 284
	Recarbonator	10.4 - 12.1	48 - 132	40 - 212	112 - 240	0	164 - 220

TABLE 11

SUMMARY OF AVERAGES FOR SELECTED CONSTITUENTS IN STATISTICAL DISTRIBUTION ANALYSIS OF POND NO. 3

(See Tables 4, 5, & 6 and Figure 16)

	Dec. 31	Jan. 7	Jan. 16
Suspended Solids Averages			
S (Shallow) mg/l	111	51	70
M (Middle) mg/l	46	61	55
L (Lower) mg/l	40	62	30
Phosphate			
Average, mg/l	38.5	28.7	40.5
Maximum, % above Avg	+ 6.5%	+ 11%	+ 6%
Minimum, % below Avg	- 4%	- 27%	- 9%
Dissolved Oxygen			
Average, mg/l	-	4.35	7.65
Maximum, % above Avg	-	+ 27%	+ 26%
Minimum, % below Avg	-	- 24%	- 22%
Dissolved Solids			
Average, mg/l	556	614	545
Maximum, % above Avg	+ 9%	+ 13%	+ 6.5%
Minimum, % below Avg	- 11%	- 29%	- 7.3%
Ammonia Nitrogen			
Average, mg/l	12.6	9.6	11.2
Maximum, % above Avg	+ 11%	+ 19%	+ 12.5%
Minimum, % below Avg	- 11%	- 21%	- 14%

TABLE 12: DATA SUMMARY -- AIR FLOTATION WITH ALUMINUM SULFATE COAGULATION AND pH ADJUSTMENT

Run No.	Alum. Sulfate mg/1	Sulfuric Acid mg/1	UCAR-149 mg/1	Chem. Costs. $/M.G.	Influent						Air Flotation Effluent						Filter Effluent(3)				
					pH	$PO_4^{=}$ mg/1	S.T.U.	Nitrogen-mg/1 NH_3	Nitrogen-mg/1 Organ.	C.O.D.	pH	$PO_4^{=}$ mg/1	S.T.U.	Nitrogen-mg/1 NH_3	Nitrogen-mg/1 Organ.	C.O.D. mg/1	$PO_4^{=}$ mg/1	S.T.U.	Nitrogen-mg/1 NH_3	Nitrogen-mg/1 Organ.	C.O.D. mg/1
Footnote-				(1)			(2)					(4)	(2)	(4)	(4)	(4)	(4)	(2)	(4)	(4)	(4)
4	240	520	-	$136.00	8.4	-	-	-	-	-	4.8	-	33	-	-	-	-	4	-	-	-
5	240	300	-	104.00	8.6	40.0	130	17.4	-	258	6.6	1.5	13	16.8	-	85	1.5	-	-	-	58
6	240	-	0.5	64.20	8.3	-	170	-	-	-	7.4	-	75	-	-	-	-	-	-	-	-
7	240	-	1.0	68.40	8.2	-	120	-	-	-	7.2	-	44	-	-	-	-	-	-	-	-
9	192	300	1.2	101.80	8.5	28	175	20.4	-	-	5.1	18	13	22.2	-	-	-	-	-	-	-
10	240	300	1.0	112.20	7.8	-	138	-	-	-	5.5	-	11	-	-	-	-	-	-	-	-
11	180	384	1.25	113.00	8.3	43	148	17.5	-	318	6.2	18	23	17.0	-	129	-	-	-	-	-
12	180	300	1.0	97.00	8.4	-	176	-	-	-	6.2	-	18	-	-	-	-	4	-	-	-
13	180	300	1.0	97.00	8.5	35	98	13.0	-	392	5.9	12	21	15.0	-	159	-	4	-	-	-
16	120	300	3.0	99.00	8.1	46.0	126	18.2	22.4	-	5.1	50.0	35	12.3	18.9	-	50.0	10	8.4	16.8	-

(1) Chemical costs are calculated per million gallons based on the delivered costs of chemicals to the Lancaster area, Table 7.

(2) Averaged values, not corrected for dilution with fresh water flow stream.

(3) Effluent was filtered through a diatomaceous earth filter or through laboratory filter paper.

(4) Averaged values, corrected by factor of 1.50 for 50% fresh water dilution stream.

TABLE 12

TABLE 13: DATA SUMMARY -- AIR FLOTATION WITH LIME COAGULATION

Run No.	Lime Dosage mg/1	Coagulant Aids		Chemical Costs per M.G.	Influent						Air Flotation Effluent						Filter Effluent
		Type	mg/1		pH	S.T.U.	PO_4* mg/1	Nitrogen NH_3	Nitrogen Organ.	C.O.D. mg/1	pH	S.T.U.	PO_4* mg/1	Nitrogen-mg/1 NH_3	Nitrogen-mg/1 Organ.	C.O.D. mg/1	PO_4* mg/1
Footnote-				(1)		(2)	(5)	(5)	(5)	(5)		(2)	(5)	(5)	(5)	(5)	(5)
15	800	NONE	-	$66.80	8.3	175	-	-	-	-	12.1	9	-	-	-	-	-
17	800	NONE	-	66.80	8.6	168	-	-	-	-	11.8	14	-	-	-	-	-
18	600	NONE	-	50.00	8.2	168	58.0	20.3	-	400	11.7	18	10.0	18.4	-	76	7.8 (3)
19	600	NONE	-	50.00	9.4	168	-	-	-	-	11.8	29	-	-	-	-	-
20	400	NONE	-	33.40	8.2	140	-	-	-	-	11.5	25	-	-	-	-	-
21	400	NONE	-	33.40	8.2	118	49.6	20.3	10.0	277	11.8	18	21.0	18.9	6.3	95	-
22	400	NONE	-	33.40	8.2	134	-	-	-	-	11.8	21	-	-	-	-	-
23	200	$FeSO_4$	100	39.60	8.3	126	-	-	-	-	10.5	120	-	-	-	-	-
24	300	NONE	-	25.00	8.2	154	47	-	-	-	11.2	65	10.0	-	-	-	0.6 (4)
25	600	NONE	-	50.00	7.8	120	-	-	-	-	11.6	29	-	-	-	-	-
26	400	$Fe_2(SO_4)_3$	25	57.15	7.8	120	-	-	-	-	11.4	32	-	-	-	-	-
		H_2SO_4 (6)	120														
27	400	$Fe_2(SO_4)_3$	12.5	45.30	7.8	126	-	-	-	-	11.2	46	-	-	-	-	-
		H_2SO_4 (6)	60														
28	300	$Fe_2(SO_4)_3$	12.5	37.00	7.9	139	-	-	-	-	10.1	91	-	-	-	-	-
		H_2SO_4 (6)	60														
29	300	$Fe_2(SO_4)_3$	12.5	37.00	7.9	152	-	-	-	-	10.4	89	--	-	-	-	-
		H_2SO_4 (6)	60														

(1) Chemical costs are calculated per million gallons based on the delivered costs of chemicals to the Lancaster area, Table 7.

(2) Averaged values, not corrected for dilution with the fresh water pressurized flow stream.

(3) Filtered through a one square foot diatomaceous earth filter.

(4) Filtered through laboratory filter paper.

(5) Averaged values, corrected by factor of 1.5 for 50% dilution with fresh water stream.

(6) Sulfuric acid was used to help dissolve Ferric sulfate prior to addition to flow stream.

TABLE 13

TABLE 14: TURBIDITY AND SOLIDS DATA FOR AIR FLOATATION TEST RUNS

Date	Air Flot. Run No.	Sampling Point	pH	Turbidity S.T.U.	Total[(1)] Solids mg/l	Suspended[(1)] Solids mg/l	Dissolved[(1)] Solids mg/l	Coagulants Used[(2)] Type	Coagulants Used[(2)] mg/l
4/6/64	5	Pond 3 water	8.6	130	636	110	526	$Al_2(SO_4)_3$	240
		Air Flot. Effl.	6.6	13	477	32	445	H_2SO_4	307
		Filtered Effl.(3)	6.6	-	626	0	626		
4/9/64	9	Pond 3 water	8.5	175	709	54	655	$Al_2(SO_4)_3$	192
		Air Flot. Effl.	5.1	13	527	15	512	UCAR-149	1.20
								H_2SO_4	300
4/22/64	11	Pond 3 water	8.3	148	717	114	603	$Al_2(SO_4)_3$	180
		Air Flot. Effl.	6.2	23	549	46	503	UCAR-149	1.25
								H_2SO_4	384
4/27/64	13	Pond 3 water	8.5	98	727	144	583	$Al_2(SO_4)_3$	180
		Air Flot. Effl.	5.9	21	559	68	491	UCAR-149	1.00
		Filtered Effl.(4)	-	4	-	-	-	H_2SO_4	300
4/28/64	16	Pond 3 water	8.1	126	640	46	594	$Al_2(SO_4)_3$	120
		Air Flot. Effl.	5.1	35	684	30	654	UCAR-149	3.0
		Filtered Effl.(4)	-	10	648	14	634	H_2SO_4	300

(1) Solids Concentrations have not been corrected for dilution by the fresh water pressurization stream.

(2) All coagulants were added to Pond 3 water prior to air flotation.

(3) Filtering was performed in the laboratory through a Scheicher & Schuell No. 589 medium filter paper.

(4) Effluent was filtered through a one square foot diatomaceous earth filter.

TABLE 15: CHEMICAL & SOLIDS DATA FOR DENSOPAC

Date	Densopac Run No.	Sampling Point	pH	Turbidity S.T.U.	Phosphate mg/1	Ammonia (as N) mg/1	C.O.D. mg/1	Suspended Solids mg/1	Dissolved Solids mg/1	Total Solids mg/1	Coagulants Used (4) Type	Coagulants Used In Densator mg/1	Coagulants Used In Recarb. mg/1	Chemical Costs (3) Per Million Gallons
3/27/64	A	Influent (1)	-	-	45	-	-	-	-	-	Lime	610	-	$51.00
		Densator Eff.	-	-	15	-	-	-	-	-				
		Recarb. Eff. (5)	-	-	6	-	-	-	-	-				
4/6/64	5	Influent	8	148	41.0	19.0	276	112	526	638	Lime	655	-	96.00
		Densator Eff.	12.4	26	3.0	19.6	96	70	709	779	SF-16	2.0	2.0	
		Recarb. Eff. (5)	10.4	32	0.0	19.6	99	74	547	621				
4/24/64	11	Influent (1)	8.9	170	36	-	338	98	647	745	Lime	655	-	92.00
		Densator Eff.	12.0	20	0	-	133	134	850	984	SF-16	2.0	-	
		Recarb. Eff. (5)	11.6	16	0	-	97	100	574	674	UCAR-149	-	2.0	
		Filter Eff. (2)	-	-	0	-	73	26	545	571				
		Filter Eff. (2)	-	-	-	-	-	50	533	603				
4/29/64	14	Influent (1)	8.3	153	48.4	21.0	-	42	649	691	Lime	655	-	93.50
		Densator Eff.	12.0	20	6.2	19.6	-	28	983	1011	SF-16	-	3.0	
		Recarb. Eff. (5)	11.7	28	4.0	18.2	-	28	528	556	WPB-25	2.0	-	
		Filter Eff. (2)	-	-	5.0	19.6	-	6	542	548				
5/19/64	18	Influent (1)	7.8	120	47	-	-	-	-	-	Lime	400	-	61.50
		Densator Eff.	11.5	35	7.0	-	-	-	-	-	$CaCl_2$	100	-	
		Recarb. Eff. (5)	10.4	-	6.8	-	-	-	-	-				
		Filter Eff. (2)	-	-	2.1	-	-	-	-	-				

(1) Influent was from oxidation Pond No. 3.

(2) Recarbonator Effluent was filtered through a one square foot diatomaceous earth filter.

(3) Costs are based on all coagulants used but exclude the cost of carbon dioxide used in the recarbonator.

(4) SF-16 is Superfloc-16 of the American Cyanamide Co., UCAR-149 is manufactured by the Union Carbide Co., WPB-25 is a non-ionic polymer of the Stein-Hall Co.

(5) Recarbonator treatment included about 1000 mg/1 of carbon dioxide; there was no way of measuring actual quantities used.

Joint PHS-LACSD Water Renovation Research Project

PART TWO

FOAM SEPARATION STUDY

Pomona, California

Malcolm L. Whitt - Project Engineer

John Zoltek, Jr. - Assistant Project Engineer

Foam Separation Study

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PROGRESS REPORT ON
FOAM SEPARATION STUDY

CHAPTER 1

SUMMARY OF FINDINGS

The most significant results obtained in the study are summarized in plotted form in Figure 1. The details and features of the pilot foamer are described in Chapter 2-c. Most of the data was obtained at a 0.5 mgd flow rate and a 5 minute residence time.

Air to water ratios ranged from 0.1 to 1.5 cubic ft of air per gallon of water. ABS and COD removal efficiencies are plotted in a wide band since removal efficiencies varied by as much as 10% for apparently identical operating conditions. The differences may be ascribed in part to normal variations in samples and analyses or to actual changes in efficiency due to daily variations in temperature and humidity or actual variations in the nature and composition of the spectrum of molecules which are detected by the COD and ABS tests.

Influent ABS values ranged from 2.7 to 6.3 mg/l and influent COD from 42 to 94 mg/l. In general, the higher the influent concentration of ABS, the better the percent removal. Whereas, the higher the influent COD concentration the poorer the percent COD removal. The latter finding was not as consistent as the former since it depended primarily on whether the increased influent COD was related to surface active or non-surface active materials.

Based on the best available information on the equivalent COD value of ABS, only about 75% of the reduction in COD can be attributed to the combined removal of ABS and suspended solids obtained in the foamer. This finding needs further checking but it supports the possibility that foaming may be a useful tertiary process for COD reduction even after the change in ABS formulation.

Careful sampling of the water near the foam-water interface did not reveal any signs of surface enrichment phenomena. A four-inch slab of Hexcell was installed below the water surface in an effort to prevent turbulence at the interface. However, the Hexcell did little to retard exchanges of water through it.

Previous laboratory studies by others have resulted in the theory that the majority of the surface active material attracted to a bubble is collected during its formation and that relatively little additional removal is obtained as the bubble travels through the water. Based on the findings of this study, there is a required minimum average contact time between the bubble and the water to achieve maximum removals of ABS as the ABS concentrations at the bubble surface and in the water come to equilibrium.

It appears that the percentage of ABS removal that can be attributed to the moment of bubble formation is relatively small.

Changing the freeboard (the distance between the original water surface and the edge of the foam overflow baffle) has a dramatic effect on the quantity of foamate as shown in Figure 1. High foamate flow means a greater foamate disposal problem and weaker ABS concentration in the foamate. Greater freeboard allowed more time for water drainage out of the foam thereby greatly reducing the quantity of foamate while slightly reducing the efficiency of removal.

CHAPTER 2

INTRODUCTION

a. Objective

The study objective was to evaluate at full scale the theoretical and physical phenomena related to foam separation techniques and apply the findings to the design of an economical, effective, prototype foamer.

The interest in foam separation as a treatment process was originally generated by the need to remove alkyl benzene sulfonate (ABS) surfactants of detergent origin from well-treated waste waters. Since laboratory studies first commenced, the detergent industry has made significant progress on the modification of the ABS molecule to a form more readily degraded in biological treatment processes. Complete conversion to the new materials is anticipated by the middle of 1965; foam separation should no longer be required for ABS removal after that time. However, during the course of laboratory studies, it has been determined that there are some organic compounds other than ABS in waste waters which are sufficiently surface active to be removed with ABS foam even though they cannot produce sufficient froth for removal on their own. These materials have not been isolated or identified but are indicated by reductions in

COD during foam separation which cannot be attributed either to the ABS or suspended solids removed.

The study, therefore, was aimed primarily at developing cost data, which could be scaled to any capacity foamer that might be required, while refuting or substantiating the possibility that foam separation may be a desirable tertiary process even after the conversion to a more readily biodegradable surfactant. It would be entirely feasible to add low concentrations of degradable surfactants prior to a foamer if a significant removal of suspended and dissolved organic matter in addition to the removal of the surfactant could be achieved. The principle is similar to adding powdered activated carbon to water for removal of organic matter but the cost is considerably less.

Foaming of primary effluents has been attempted unsuccessfully in the past; the high suspended solids content seems to interfere with stable foam formation. Recent laboratory tests by the PHS have indicated that it may be possible to accomplish significant removals of BOD and suspended solids from primary effluents using coagulant aids and foam separation. In some cases the additional treatment by foaming may obviate the need for more elaborate and expensive secondary treatment processes that are not justified by the method of effluent disposal.

b. Location

The study was conducted at the Sanitation District No. 21 Water Reclamation Plant at 395 South Roselawn Avenue, Pomona, California. The plant receives about 4.5 mgd of the 9 mgd of waste water originating in the cities of Pomona, Claremont and La Verne and some unincorporated area in Sanitation District 21. Waste water not treated at the plant is bypassed around the plant and conveyed in the District 21 Outfall to a jointly operated primary treatment plant and ocean outfall disposal system near the Palos Verdes Hills owned by the County Sanitation Districts. All solids wasted from the Water Reclamation Plant are discharged to the trunk sewer; no sludge digestion occurs at the plant. Treated effluent is either sold to an irrigation water company or discharged to San Jose Creek for percolation into downstream ground water basins.

The plant was originally constructed in 1925 as a standard rate activated sludge plant but in 1962 modifications were completed to permit operation of the old structures by the contact stabilization process. Plant flows vary from 1 to 6 mgd and average about 4.5 mgd. Influent COD averaged 700 mg/l, primary effluent COD averaged 300 mg/l, and secondary effluent COD averaged 50 mg/l during the course of the study. Return sludge is aerated for approximately 8 hours prior to mixing with primary effluent and the mixed liquor is aerated for 2.8 hours. The plant achieves nitrification with the nitrate nitrogen (NO_3 as N) ranging from 2 to 10 mg/l in the effluent and averaging about 8 mg/l.

c. Project Staff

The project staff was comprised of:

Mr. Malcolm L. Whitt, Project Engineer, C.S.D.
Mr. John Zoltek, Jr.,Asst. Project Engineer, P.H.S.
Mr. Nathaniel C. Anderson, Chemist, C.S.D.

Mr. Whitt was particularly responsible for supervising construction of the pilot foamer.

Mr. Zoltek operated the foamer and plotted and analyzed the data obtained.

Mr. Anderson performed the laboratory analyses for the foamer and maintained enough plant operating data to provide control on the quality of the water produced at the reclamation plant.

The initial conceptual design of the foamer was by Dr. David G. Stephan, Deputy Chief of the USPHS Advanced Waste Treatment Research Program.

Construction of the foamer, with related piping and modifications, was performed by Sanitation District Maintenance personnel under the direction of Mr. Jim Mikkelsen, foreman.

General supervision and planning was the responsibility of Mr. Franklin D. Dryden, Project Coordinator. Overall joint responsibility for the project rested on Mr. Gerald N. McDermott, Project Officer from the USPHS Advanced Waste

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Treatment Research Section and Mr. John D. Parkhurst, Project Director and Chief Engineer and General Manager of the County Sanitation Districts of Los Angeles County.

d. Equipment

The foamer was designed, constructed and placed in operation between September and November, 1963. Its capacity of 1,625 gallons was based on a design flow of 0.5 mgd at a 5 minute residence time. The foamer is 15 feet long, 3.5 feet wide, and 9 feet high, constructed of steel and situated on a 6' wide by 32' long concrete slab with a 1% slope. (Figure 2) The side-wall and top are removable and there is an access hole in one end. A lifting bar supported at the ends of the tank permits lifting the top without the aid of a crane.

Pomona Water Reclamation Plant secondary effluent is pumped to the unit through a manual throttling valve and 4" propellor meter with flow indicator and totalizer. Influent is discharged toward the bottom of the head end of the tank. Effluent is discharged from the bottom of the discharge end through a stand-pipe with removable three-inch rings for water level control.

Air is supplied from the main plant air blower at about 7 psi, and is manually throttled through two valve and orifice meter installations to air headers at two levels within the tank. The use of three sizes of orifices and 30" manometer for each air meter allows accurate control over a wide range of air rates. Each row of air headers contained about fifty 3" diameter by 2' long saran wrapped air diffusers as manufactured by the Chicago Pump Company. One row was placed at the bottom of the tank and the other row was in the middle, at the top or totally removed at different times during the year.

Foam removal was accomplished by extrusion of the foam over a baffle which extended the full length of the tank. The foam was liquified by use of a high pressure spray system operating on recycled foamate. A 35 gpm 1 hp Peerless centrifugal pump took suction from an intercepting well in the bottom of the foamate discharge line and recycled it through 18 - 1/8" diam. orifice spray nozzles (Spraying System nozzle No. 1/8 K18) onto the foam. Only a small volume of foamate was required to fill the spray system; all subsequent foamate was discharged to the adjacent channel. Foamate flows were determined by timing the filling of a 35 gallon oil drum.

Sampling taps were included on the influent and effluent lines and at 8 vertical locations for 5 cross-sections along the length of the tank. Glass portholes were provided at five locations to permit observation of a diffuser, the foam-water interface, and the foam produced at both ends of the tank.

The laboratory was small, noisy, and cramped, and it vibrated with the rest of the blower building in which it was housed. Weighing was, therefore, difficult and inherently inaccurate. However, there was sufficient room to set up 12 COD's at a time and a Bausch & Lomb Spectronic-20 colorimeter was used in most of the ABS tests performed by the "Hellige" method.

e. Testing and Sampling Procedures

Only two analyses were run on a regular basis; COD and ABS. The COD Test used in Pomona is similar to the one described in "Standard Methods" with the exception that 0.7 gm of silver sulfate (Ag_2SO_4) is used as a catalyst and 1 gm of mercuric sulfate is used to complex chloride interference.

The ABS Test used in Pomona is a modified version of the "Hellige" method. The normal "Hellige" method consists of shaking the sample in a plastic bottle containing 15 ml of methylene blue solution and 10 ml of chloroform. The chloroform phase is transfered to a color comparator vial and the ABS concentration is obtained by visual comparison with calibrated colored discs. The modification used in this study involved reading the % transmittance of the chloroform solution on a Bausch & Lomb Spectronic-20 colorimeter. To further reduce error by cancelling out interferences a standard curve of percent transmittance versus the ABS concentration found by the standard method was plotted on logarithmic paper. The line of best fit provided a maximum spread of 0.10 mg/l for 90% of the points. The test can be used on a 50 ml sample for a range of 0.6 to 1.5 mg/l of ABS without any dilution. The test required approximately 5 minutes per sample and duplicates were always run and averaged.

Prior to collecting samples, the foamer operating conditions were set and the foamer allowed to stabilize through 5 residence times (25 minutes under normal operation.) Two-liter samples were then collected slowly over a 5 to 10 minute period; the samples were stirred and 500 ml portions were separated for the tests. The chemist was able to run 24 ABS samples and 12 COD samples in an 8-hour day with an hour to spare for cleanup or emergencies.

CHAPTER 3

DISCUSSION OF RESULTS

a. General

The results obtained in the study cover a wide range of operating conditions and several variations of the basic foamer components. Unfortunately the effort to evaluate a large number of variables in the time available prevented developing any significant number of runs for a single set of operating conditions. In most cases only a single run of 5 or 6 points is available. A slight error in sampling or analysis, or an unusual change in influent concentrations during a run may well have resulted in unusual deviations of single points which materially affect the shape or range of a curve.

In addition, the study period only extended between November and May or essentially through the winter months. There is some previous experience at the Whittier Narrows Water Reclamation Plant foamer to indicate that in the summer, the activated sludge process is more effective on ABS and that even the foaming process may be affected by viscosity, surface tension, or vapor pressure variations which are temperature related.

The data have been plotted using the ratio of cubic feet of air used per gallon of influent to the tank as the basic parameter. Removals of C.O.D. and ABS in percent, and the percent the foamate flow is of the influent flow, provide the primary standards of comparison. No attempt has been made to correct the air-water ratio to account for excessive foamate flows. However, it should be noted that a 20% foamate flow would result in an effective 10% increase in the air-water ratio; at low rates of foamate flow the difference is negligible.

The dimensions of cubic feet per gallon for the air-water ratio were chosen for convenience in place of a dimensionless value since in actual practice water flows are readily available in "gpm" and air flows are conventionally measured in "cfm." The figures used herein can easily be converted to their dimensionless equivalents by multiplying the air-water ratio's by 7.48.

b. Variations in Results

In April, several runs were performed under identical operating conditions on different days in order to evaluate the reproducibility of results when processing waste water from the same source at intervals over a period of time. Five runs were performed with bottom air and a 5-minute detention time, of which three used a 1.5 ft freeboard and two used a 2.5 ft freeboard. The Hexcell slab was in place in both cases, but is considered to have no effect on the process. The results are presented in Figures 22 and 23.

For the two runs at 2.5 ft freeboard the results are very consistent for foamate flow and ABS. The C.O.D.'s correlated closely at three points and were spread by 4 to 10% in the other three.

For the three runs at 1.5 feet of freeboard the correlation was less satisfactory but the following differences should be noted:

The run on April 2 had both the lowest influent ABS concentrations (3.0 to 3.85 mg/l) and lowest COD concentrations (average 48.5 mg/l). The ABS percent removals for April 8 and 14 were quite close and so were the influent ABS concentrations for those two days (3.75 to 5.0 mg/l on the 8th and 3.9 to 4.8 mg/l on the 14th.) COD values were highest on the 8th (average 54.7 mg/l) and the COD on the 14th (average 51.5 mg/l) was essentially half-way between the other two.

However, foamate flows for air rates in excess of 0.5 cu.ft/gal were consistently and appreciably highest on the 14th and lowest on the 2nd. In addition, C.O.D. removals on the 14th were not only greater in % than removals on the 8th but in actual amounts also. One can only conclude in the light of almost identical ABS concentrations and removals on the 8th and 14th, that the increased foamate flow and C.O.D. removal on the 14th was because a greater percentage of the non-ABS COD present on the 14th was surface active and therefore was removed with the increased foam it helped to create. Conversely one could speculate that the somewhat higher C.O.D. on the 8th was comprised of suspended matter which inhibited foaming.

It can reasonably be argued that both the composition and the quantity of C.O.D. in the plant effluent are more variable than the composition and quantity of the ABS. Reductions in ABS through the plant are relatively unaffected by changes in plant operation which may materially affect the C.O.D. For example, a condition of rising sludge which occasionally occurred during the study, could increase plant effluent C.O.D. from about 45 to 80 mg/l without noticeably affecting ABS concentrations. Under such conditions the COD of the suspended solids easily represented 60% of the total C.O.D. compared to a more normal 30%. Even though C.O.D. removal in mg/l would be higher when suspended solids were high, the percent removal generally decreased. A brief investigation of the relationship between suspended matter and COD removals in foam separation is reported in the last subsection of this chapter.

Since COD is composed of a wide spectrum of suspended and dissolved material it is not only more easily affected by plant operation than ABS but is subject to more fluctuation at its source. It was observed during the study that ABS values followed a consistent daily trend whereas COD values did not. But in addition to being inconsistent in composition, the COD concentration has little to do with its own removal. Since the foam of ABS origin provided the primary mechanism for COD removal, the extent of COD removal is closely related to the original ABS concentration and relatively unrelated to its own concentration.

The ABS also comprises a significant percentage of the COD removed. Tests performed at the Robert A. Taft Sanitary Engineering Center indicate that 1 mg/l of ABS exerts a COD of about 2 mg/l. The removal of 2.0 mg/l of ABS by foaming would equal a 10% reduction in the COD of a 40 mg/l COD solution but only a 5.0% reduction in an 80 mg/l solution. In relation to a total COD reduction of 30 to 40%, these differences are significant.

c. Bottom Air

Operation of the foamer with bottom air only gave the best percent removal of ABS and COD and the highest foamate flow for a given air rate and freeboard. Both findings can be related to the fact that by releasing all the air at the bottom of the tank, a greater volume of air was in the water at any one time and the average contact time between air and water was longer than for any other arrangement. The greater volume of air in the tank and the correspondingly greater displacement of water results in an effective decrease in the freeboard coupled with less drainage of the foam and greater foamate carryover. This alone would account for slightly higher removals per unit of air, but in addition the increased contact time in the tank helps to assure that each bubble will have reached equilibrium with the solution through which it is rising before it escapes.

In some of the runs with bottom air, the tank still contained the upper row of air diffusers or the 4" slab of Hexcell. It is not believed that these features affected the results because of the high degree of turbulence within the tank. More data was obtained with bottom air only than with any other configuration.

It will be noted in Figure 3 that the improvement in relative removal efficiency for both ABS and COD does not progress uniformly with freeboard depth. By the curves obtained, the best combination of low foamate flow coupled with highes removal of ABS and COD would occur with a 1.5 ft freeboard at about 0.5 cu.ft./gal of air. Though it may not be possible to improve removals by lowering freeboards below 1.5 ft, the recorded decreases in efficiency are probably related to the normal variations which result from performing runs on different days as described previously.

There are presented in Figure 4 the results of bottom air only runs at 2.5 minute and 10 minute detention times. Since air rates are proportional to flow rates but the average time for a bubble to get from the diffuser to the water surface is approximately constant regardless of air or water flow rates, at a 2.5 minute detention time there is about four times the volume of air in the tank for any equivalent air-water ratio as for a 10 minute detention time. This fact is illustrated by the excessive quantity of foamate for the 2.5 minute detention time in spite of a 2.5 ft freeboard compared to the less than 1% foamate flow for the 10 minute detention time with only a 1.5 ft freeboard. However, removal efficiencies at the 10 minute detention time were significantly below those at the 2.5 minute detention time until air rates reached 0.7 cu.ft./gal after which it was better. At very low air rates, the volume of foam generated was probably quite small and even at intermediate air rates the low percent foamate flow indicated good drainage was occurring with corresponding return of surfactants to the water. Only at the higher air rates was there enough air per square foot of tank cross section to generate a stable foam, allow it to drain and still achieve maximum removals.

d. Middle Air

Performance of the tank with air diffused only at the middle depth (Figure 5) was similar but less efficient than with air only at the bottom. The shallower depth resulted in a smaller volume of air displacing the water; the effective water depth was therefore not as great which is an explanation for the lower foamate flows. Turbulence in the tank certainly resulted in bubble movement below the row of diffusers and even to the bottom of the tank but the average contact time between bubbles and water was reduced. At air rates of 1.4 cu.ft./gal, foamate flows increased appreciably and COD removals exceeded the results obtained with bottom air at less than 1.0 cu.ft./gal air rates.

One advantage of using air diffusion at some level above the bottom is that, if efficiencies are not reduced more than 10 to 15%, it may be possible to effect a cost savings by supplying somewhat more air at reduced pressure. However, if the air is being supplied from the main plant blowers at a normal pressure of 6 or 7 psi, there can be no saving in power by discharging air at shallower depths.

e. Top Air

While the tank was arranged for the surface enrichment studies with a row of diffusers just below the water surface, a run was conducted with top air only just for the sake of completeness. (Figure 6) Even with this arrangement there was enough turbulence in the tank to effect some removal of ABS and COD although percent removals were only about half of those obtained in other runs. It is suspected that the removal that did occur was primarily accomplished by bubbles which traveled down into the main volume of the tank before escaping from the water surface. This approach should certainly not be considered in practical foamer designs.

f. Split Air

Split air runs were of two types: (1) with the air divided between the bottom and the middle and (2) with the air divided between the bottom and a row of diffusers just under the water surface. The data for a 50-50 split of air between the bottom and the middle is presented in Figure 8. ABS removal was less for the split air than for a comparable air rate in the bottom diffusers only. Up to approximately 0.6 cu.ft./gal the removal for the split air run closely matched the run for middle air only but at higher air rates the split air was more efficient. Foamate flow was only slightly reduced and might even have been a little greater compared to bottom air only at air rates under 0.5 cu.ft./gal.

There seems to be no economic advantage to splitting the air between a bottom and middle level. For example, a 60% removal of ABS was effected with 0.58 c.ft./gal of air divided equally between the bottom and middle row of diffusers. In Figure 3 it can be seen that an equivalent ABS removal was accomplished with 0.45 cu.ft./gal of bottom air only. If it is assumed that the cost of air is approximately proportional to its discharge pressure and thus to its depth, then the additional 0.16 cu.ft./gal of air supplied to the bottom costs about the same as the 0.29 cu.ft./gal of air previously diffused in the middle. With no substantial difference in cost for equivalent removal, it should be noted that there is a capital savings involved with using only one level of diffusers and the foamate flow for 0.45 cu.ft./gal of bottom air is only 1.6% compared to a 4% foamate flow for 0.58 cu.ft./gal of split air.

Much effort went into attempts to enhance the removal efficiency by distribution of air both at the bottom and near the top as described more fully in section "g" on "Surface Enrichment Phenomena." Data on top and bottom split air runs are presented in Figures 8, 17 and 21. Figure 8 presents the data for runs where the top diffusers received 10, 20, 30 and 75 percent of the total air and Figure 22 presents data for 50-50 splits of air before and after the installation of Hexcell. In general, the less aid going to the top diffusers the better the percent removals and at up to 50% top air there was no apparent increase in removals over that previously accomplished by the portion of the air supplying the bottom diffusers alone. In addition foamate flows were up to 5 times as great as for the equivalent percent removal using bottom air alone.

g. Surface Enrichment Phenomena

One of the basic theories to come out of laboratory work on ABS removal was the concept of surface enrichment or the belief that surface active materials were concentrated near the foam-water interface by the combined action of rising bubbles depositing materials from within the tank near the water surface and drainage back into the water from the foam. In addition it was postulated that most of the surface active material removed by a bubble became attached as the bubble was formed and little if any material was added as the bubble traveled through the liquid. The maximum concentration of surface active material on the bubble surface could therefore be obtained by increasing the concentration of the solution in which the bubble was formed. Based on these theories, the optimum foamer design would feature a quiescent zone just under the foam-water interface, air distribution at the bottom of the tank to move material on bubble surfaces into the zone, and the formation of additional bubbles in the "enriched" quiescent zone by a separate level of diffusers.

Placing diffusers at the bottom and near the top of the tank was no particular problem, but a truly quiescent zone was never obtained in spite of a 4" slab of Hexcel installed under the top diffusers as described in section 1. Turbulence was random and severe at higher air rates but profile sampling down the length of the tank indicated a continuous longitudinal gradient (Figures 16 to 19) in spite of relatively thorough mixing at any particular transverse cross section. Careful sampling of the water at the foam-water interface with a special sampling tube showed it to have essentially the same concentration as elsewhere in the same vertical cross section. (Figure 19) There was neither an increase as theory surmised or a decrease as might be expected if there was any significant removal accomplished in this area by the bubbles formed there. If most of the removal was truly accomplished as the bubbles are formed and the concentration is the same at both levels of diffusers, then each level should accomplish about 50% of the total removal. Such was not the case as the addition of the upper air had little or no effect at a 50-50 air split and produced a significant reduction in total efficiency when the top diffusers received 75% of the air. Whatever removal could be credited to the top air could easily be ascribed to the bubbles which were carried by tank turbulence below the diffusers and through the tank volume before reaching the foam-water interface. On the other hand, since some ABS was removed using top air only it can be argued that it occurred as the bubbles were formed. Certainly some surface active materials do form on the bubble-water interface as the bubbles are formed, but even if a truly quiescent zone could be accomplished it does not appear that the possible benefits of forming bubbles near the surface could outweigh the disadvantage of decreased contact time.

It is the conclusion of this study, that there may be no such phenomena as "surface enrichment" and that there is a minimum amount of contact time required between a bubble and its environment to establish an equilibrium between the concentrations of surfactants on the bubble surface and in the water through which it is traveling. Greater depth of diffusion will help assure that a maximum percentage of the bubbles formed receive the minimum contact time. It is not yet certain whether there is any advantage in being deeper than the 6 ft depth used during this study but it should be investigated.

h. Freeboard

Changing the freeboard seems to affect removal efficiencies only to a slight degree. Not enough runs were made under identical conditions to get an average removal curve for any set of freeboards for comparison. Based on the available data, it appears that there are greater removals at the smaller freeboards, as expected. The greater freeboards allow more time for material to drain back into the water before the foam goes over the weir.

Changes in freeboard primarily result in wide variations in the amounts of foamate carried over the weir. There is some variation in foamate flow under identical operating conditions due to daily fluctuations in plant effluent characteristics. However, this effect is small compared to the changes observed by changing the freeboard from 0.5 to 2.5 ft. (Figures 9 to 13.) For example, with a 0.7 cu.ft./gal air rate and a 1.0 ft freeboard, about 40% of the influent is carried away with the foam whereas at the same air rate and a 2.5 ft freeboard the foamate only amounts to 1% of the process flow. As previously mentioned the curves have not been corrected to show true air-water ratios as a result of excessive foamate flows and a 40% foamate flow would represent an effective 20% increase in the air rate per gallon.

For most removal processes, the cost of foamate handling and disposal increases with increasing quantities of foamate. It is therefore desirable to maintain the minimum foamate flow commensurate with required removal efficiencies and unit capital and operating costs.

i. Detention Time

Increasing the detention time from 5.0 minutes to 10 minutes and reducing it to 2.5 minutes did not have a significant effect on removal efficiencies at equivalent air-water ratios. (Figures 14 and 15). Both ABS percent removal curves fall within the range for bottom air and a 5 minute detention time presented in Figure 3. Although better removal efficiencies might have been expected for the longer detention time, the 2.5 minute curve is at the high side of the range and the 10 minute curve is at the low side. This finding is more understandable if it is realized that there is approximately 4 times the volume of air in the tank at 2.5 minutes detention time as compared to the 10 minute run, and a corresponding reduction in the effective freeboard. This fact is further illustrated by the excessive volumes of foamate produced at relatively low unit air rates at the 2.5 minute detention time and the unusually dry foam produced even at high air rates and good ABS removal during the 10 minute run.

It should be noted that the removal curve for ABS during the 10 minute run appears to still be rising at 0.9 cu.ft./gal and 75% removal and that even better removals might be accomplished at higher air rates. However, a greater capital investment and more air per gallon is required to achieve maximum removals at minimum foamate flow. Only if the cost of foamate disposal offset the increased cost of limiting foamate production at desired efficiencies would increased detention times be justified.

Generally, the foamate volume at the 2.5 minute detention time would be considered excessive at desired removals; even the initial freeboard of 2.5 ft did not prevent excessive carry-over. In the 6' deep tank used for this study, the 5.0 minute detention time seemed to provide adequate removal efficiency at relatively low foamate flows and an air rate of 0.6 cu.ft./gal. If higher removal efficiencies are desired it would be recommended that the air rate per square ft of floor area be kept constant and the length of tank and the detention time be increased.

j. Profile Runs

Profiles of vertical and longitudinal distributions of ABS and COD during several variations in tank operation (Figures 16 to 19) were obtained by taking grab samples from faucet taps along the side of the tank. Influent and effluent samples were composited in the usual manner.

The profiles reveal good vertical uniformity at cross sections near the head, center and end of the tank, and, in spite of the turbulence, a relatively consistent decrease in concentrations longitudinally. The high degree of turbulence is illustrated by the consistent and abrupt drop in ABS concentrations between the influent pipe and the first row of sampling taps 6" inside the tank end wall. The addition of the 4" slab of Hexcel (Figures 18 and 19) did little if anything to change the pattern. It is possible that it contributed to the slight decrease in ABS concentrations near the water surface at the head end of the tank, but the reduction detected was small and occurred both with split air and bottom air only.

k. Foam Collapse

Early in the study, foam collapse and liquefaction was attempted with a "cyclone separator" constructed from a 55 gallon oil drum. Partial liquefaction was achieved when foamer air rates reached or exceeded 0.5 cu.ft./gal, but the cyclone was totally unsuccessful at lower air rates. After several attempts at modification failed to improve its performance, the cyclone was abandoned. It might have sufficed for a small range of air rates, but it is doubtful that it could ever have coped with the wide range of conditions under study.

Water sprays for collapsing foam had previously been installed successfully at the Whittier Narrows Water Reclamation Plant foamer, and so a spray system was installed on the pilot foamer. At the suggestion of P.H.S. engineers, liquefied foamate was recycled to provide the spray water and this system worked successfully at all flow ranges investigated in the remainder of the study.

In Figure 20, the relationship between foamate flow and freeboard is plotted for five air rates and their corresponding "expansion velocities" based on a 5-minute detention time. Expansion velocity in ft per minute is defined as the air rate in cfm divided by the horizontal cross-sectional area of the foamer in sq. ft.

1. Hexcel

Early observations of the original foamer revealed that the 14 transverse baffles at 1' centers in the upper 2.5 ft of the tank were ineffective in preventing turbulence near the water surface. In line with the interest in "surface enrichment" phenomena, it was theorized that a layer of material with small holes in it placed just under the transverse baffles might create the desired quiescent zone by limiting the movement of bubbles to the upward direction only. The material selected was 4" thick, had a honeycomb structure comprised of 1/4" nominal diameter hexagonal tubes, was fabricated of light weight aluminum and was called "Hexcel."

Unfortunately, the rising columns of air were less random than previously believed and where they reached the Hexcel they passed through without hindrance and moved water with them. Other points on the Hexcel surface were not directly over rising bubble columns and water passed just as easily downward through those holes. It is believed that even covering a large percentage of the open area could not have fully prevented this turbulent pumping and, owing to lack of time, it was never attempted. By all the tests available including dye injection and careful sampling above and below the Hexcel slab, the Hexcel had no effect on removal efficiencies or foamate production. (Figures 21 to 23.)

m. Dye Studies

The dye used in the study was fluorescein. Two heaping teaspoons of the powder introduced into the inlet line were enough to produce a measurable effluent absorbance reading on the Spectronic 20.

The dye concentrations for various time intervals after dye injection are plotted in Figures 25 thru 28 at 2.5, 5 and 10 minute detention times and air rates of zero, 0.2, 0.5 and 1 cu.ft./gal. The fastest short-circuiting occurred with zero air. At all air rates the internal mixing appreciably increased the time required for the peak concentration to reach the tank discharge. Peak concentrations generally were discharged in about half the theoretical detention time but the time for 50% of the dye to have passed through on a mass balance of the dye area was quite close to the calculated detention time.

In Figure 28, dye distribution curves are plotted for two runs performed under identical operating conditions of a 5 minute detention time, 0.5 cu.ft./gal air rate, 1.5 ft. freeboard, and the Hexcel in place. In the first run, the dye was injected into the influent pipe, and in the second, it was injected above the Hexcel one-third the way down the tank. The two curves are almost identical, which supports the finding that the Hexcel did not provide a quiescent zone near the water surface and that the turbulence was great throughout the tank. If there had been a quiescent upper zone, the dye would have diffused slowly into the main volume of the tank and resulted in a longer flat-topped curve rather than the sharply peaked curve shown.

n. Suspended Solids - COD Data

The COD of the suspended solids (SS - COD) in the water, was determined by filtering the sample through a Gooch crucible containing a fiberglass filter pad (Hurlbut glass fiber filter paper #934-AH), dropping the pad into a COD flask with some distilled water, and relating the resulting COD to the original volume of filtered water. Total COD and filtrate COD were also determined and the sum of the SS_COD determined by this method and the filtrate COD were almost exactly equal to the total COD. Blanks with the fiberglass filter pad in distilled water had COD values of 0.2 to 0.7 mg/l.

A brief series of laboratory analyses were conducted to determine the COD of ABS. There was no consistency in the results and ratios ranged from 1.4 to 3.9 mg of COD per mg of ABS. Since the results were so varied, it was decided to use the ratio of 2 mg COD per 1 mg ABS developed at the

Robert A. Taft Sanitary Engineering Center. In this manner, it was determined that the sum of the ABS and Suspended Solids COD's equaled about 75% of the COD removed by the foamer. (Figure 29 and Table 1.) Within the range of reliability of the COD to ABS ratio, it was therefore concluded that approximately 25% of the COD removed by foaming is neither due to ABS nor suspended solids removal.

CHAPTER 4

COSTS

a. Unit Cost Curve

The unit cost of construction, operation and maintenance of a 0.5, 1.0, and 10 mgd foamer of the approximate design used in this study has been plotted in relation to desired removal efficiencies in Figure 31. Operating costs do not include the cost of foam disposal since methods of foam disposal have not been under investigation in this study. Unit costs decrease as the size of the installation increases. For 64% removal of ABS, the cost per mgd is estimated to be $12.46, $10.75 and $9.22 respectively for a 0.5, 1.0, and 10 mgd foamer. It is anticipated that the coming year's study will allow amplification of the cost curves and the inclusion of foamate flow and detention time as variables.

b. Capital Cost

The capital cost has been based on a 0.5 mgd foamer of the design used in this study with bottom air only and a 5 minute detention time. The capital cost might vary per mgd if different detention times were selected in relation to a specific goal for removal efficiency. However, since most of the work for this study was conducted at a 5 minute detention time, there is not enough information available on which to base other curves. The estimated capital cost of

$5,000 for a 0.5 mgd foamer, amortized at 5% interest over a 10 year life results in a uniform $3.60 per million gallons at all removal efficiencies.

c. Operation and Maintenance

The major operating cost is for power for the air supply at an assumed blower discharge pressure of 3 psi. The air requirement is related to efficiency of COD and ABS removal at a 5 minute detention time by the data of Figure 1. In order to convert the air requirement to a cost figure it was assumed that it requires 3 KWH per 100 cfm of air and that power costs 1 cent per KWH plus a 66% connected load charge for a 0.5 mgd plant, 33% connected load charge for a 1 mgd plant and no connected load charge for a 10 mgd plant. The total cost comes to $1.20 per 100 cfm per day for a 0.5 mgd, $0.96 per 100 cfm per day for a 1 mgd plant and $0.72 per 100 cfm per day for a 10 mgd plant.

Foamate loss also comprised a portion of the operating cost. In a water reclamation plant, the value of the foamate is the cost of producing the water through the main plant since the water would normally be sold at cost. The value used in Pomona was $40.00 per million gallons; at a 2.5 ft freeboard this amounted to only $2.28 per mgd at a 1.0 cu.ft./gal air rate. If the water is not sold or even produced for reuse, the cost of foamate disposal still represents an increasing cost with increasing removals and air rates. Since methods of foam disposal were not evaluated, such costs are not included in the cost curve.

Maintenance costs include all labor costs with respect to operating the unit, repacking pumps, cleaning foam sprays, replacing broken diffusers, painting, and daily housekeeping. This cost has been assumed to equal about two man-days per month, or 200 hours per year, regardless of removal efficiency. At $3.25 per hour the maintenance cost of $650.00 per year equals $1.80 per day for the 0.5 mgd unit. This cost does not increase proportionately with the capacity of the unit but may reach $3.60 per day for a 10 mgd plant based on experience with the foamer at the Whittier Narrows Water Reclamation Plant. Because the maintenance cost is not proportional, it was not possible to provide a single curve of unit cost versus removal efficiency.

GLOSSARY

Freeboard - The distance from the water surface to the overflow weir under static conditions.

Bottom air - Air introduced into diffusors that are 0.25 ft from the bottom of the tank.

Middle air - Air introduced into diffusors that are 2.5 ft from the bottom of the tank.

Top air - Air introduced into diffusors that are 5.0 ft from the bottom of the tank.

Split air - Air introduced into two levels of diffusors.

Plant - Pomona Water Reclamation Plant

Foamer - Foam separation device used in this research.

Expansion velocity - Air flow in cfm divided by horizontal cross-sectional area of tank.

Foam - Foam before it is collapsed.

Foamate - Collapsed foam.

COD - Chemical oxygen demand.

ABS - Alkyl benzene sulfonate, the surface active agent in most household detergents.

SS - Suspended solids

Hexcel - Trade-name of expanded aluminum honey-comb.

Residence time - The actual time the foamer was in operation divided by the detention time. Example: 20 minute run, 5 minute detention, equals 4 residence times.

Figure 1

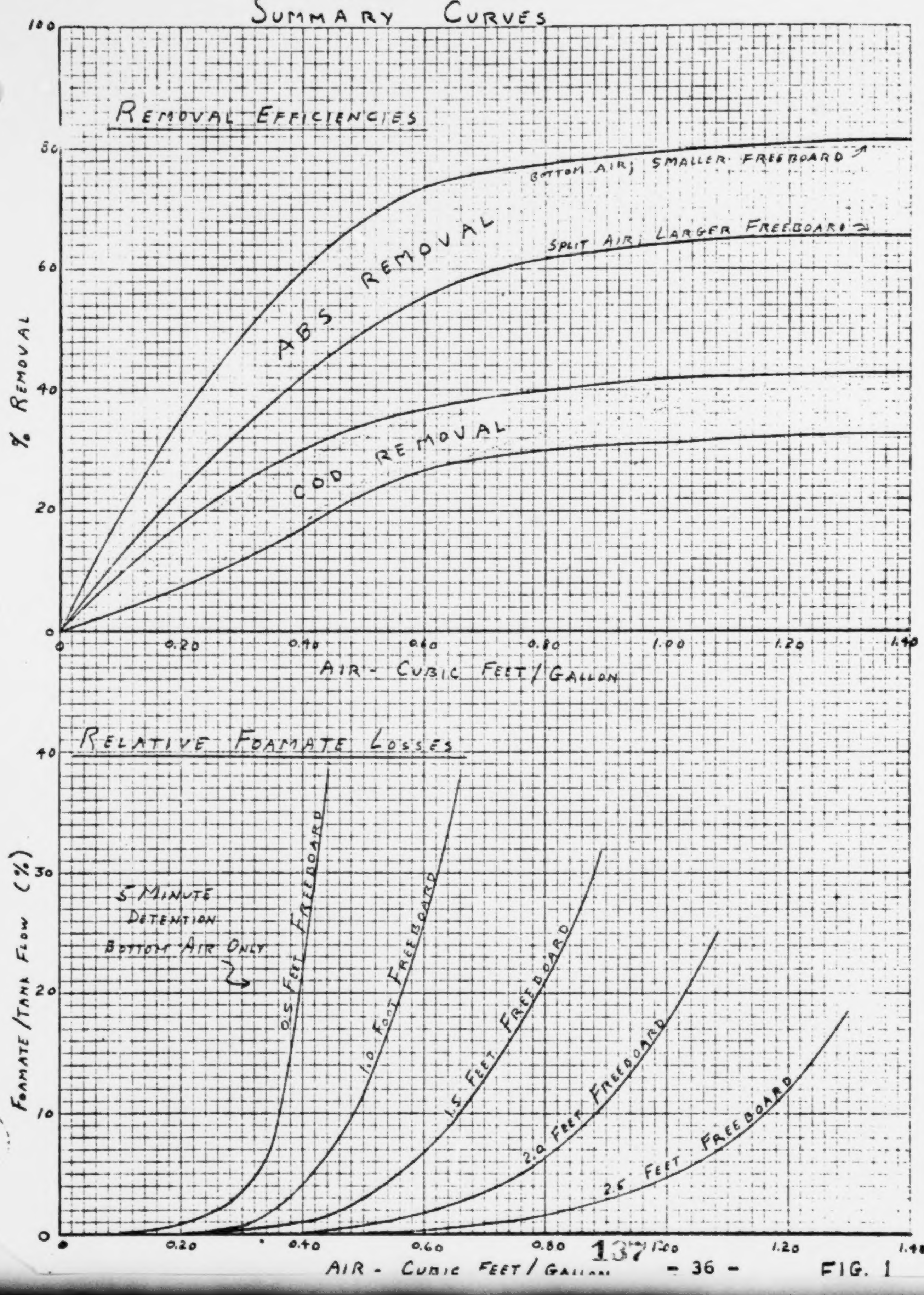

FIG. 1

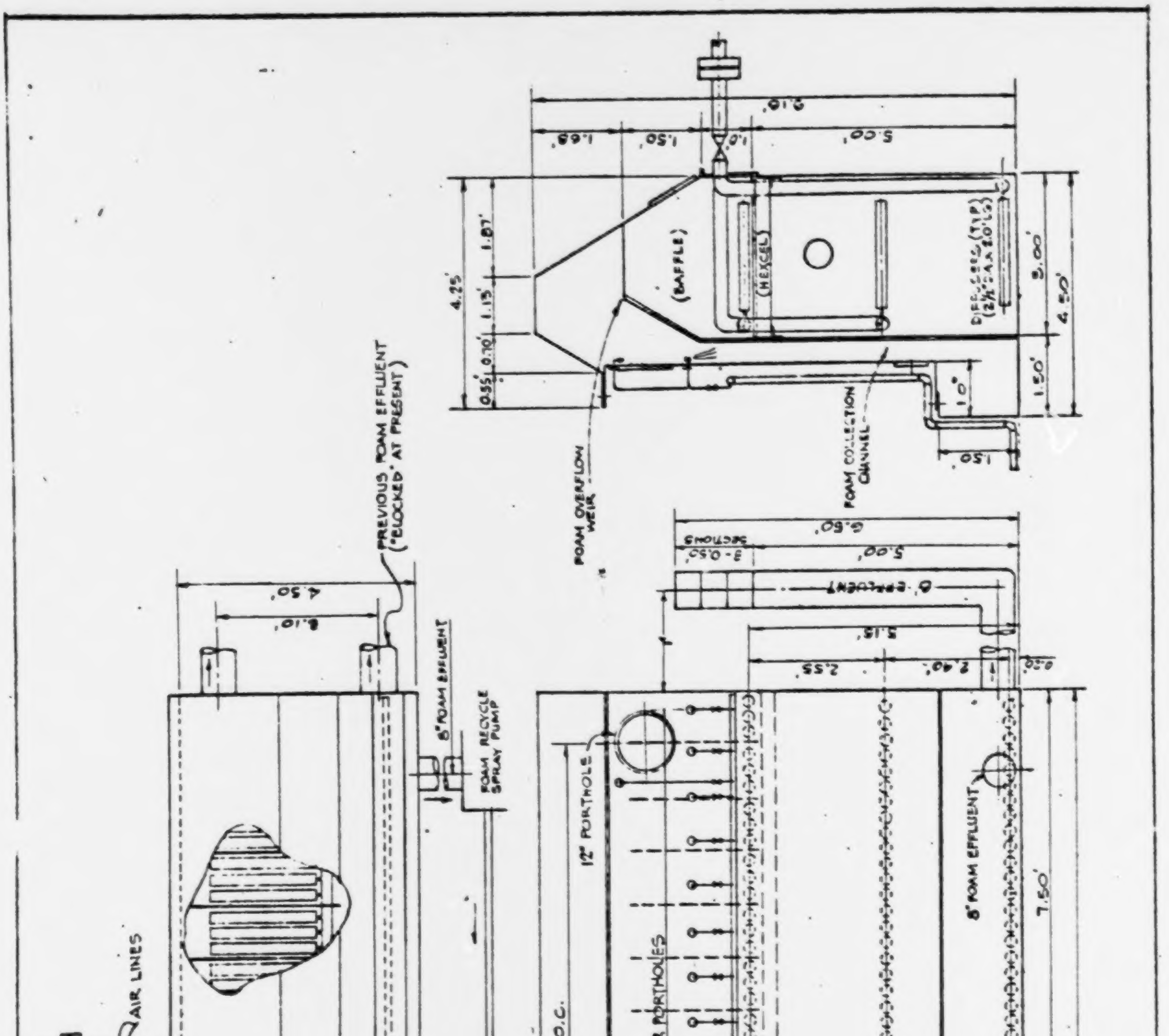
6" INFLUENT
8" PORTHOLE
10" PORTHOLE
AIR LINES
10'
PREVIOUS FOAM EFFLUENT ("BLOCKED" AT PRESENT)
8" FOAM EFFLUENT
FOAM RECYCLE SPRAY PUMP
8.10'
4.50'
14 - BAFFLES @ 12" O.C.
12" PORTHOLE
18 - SPRAYS, 10" O.D., + 2 FOR PORTHOLES
TOP ROW DIFFUSERS (45) 3 - EA. BAY
MIDDLE ROW DIFFUSERS (50)
BOTTOM ROW DIFFUSERS (50)
6" PORTHOLE
8" FOAM EFFLUENT
6" INFLUENT
6" PORTHOLE
3.9'
7.50'
9.10'
7.50'
7.50'
15.00'
0.20'
2.40'
2.55'
5.15'
8" EFFLUENT
3 - 0.50' SECTIONS
5.00'
6.60'
FOAM OVERFLOW WEIR
FOAM COLLECTION CHANNEL
(BAFFLE)
(HEXCEL)
DIFFUSERS (TYP) (2½" DIA × 20" L'G)
0.55'
0.70'
1.15'
1.87'
4.25'
1.50'
1.0'
3.00'
4.50'
1.68'
1.50'
1.0'
5.00'
9.10'

FIGURE 2 - FOAMER DESIGN

Figure 3 - Bottom Air

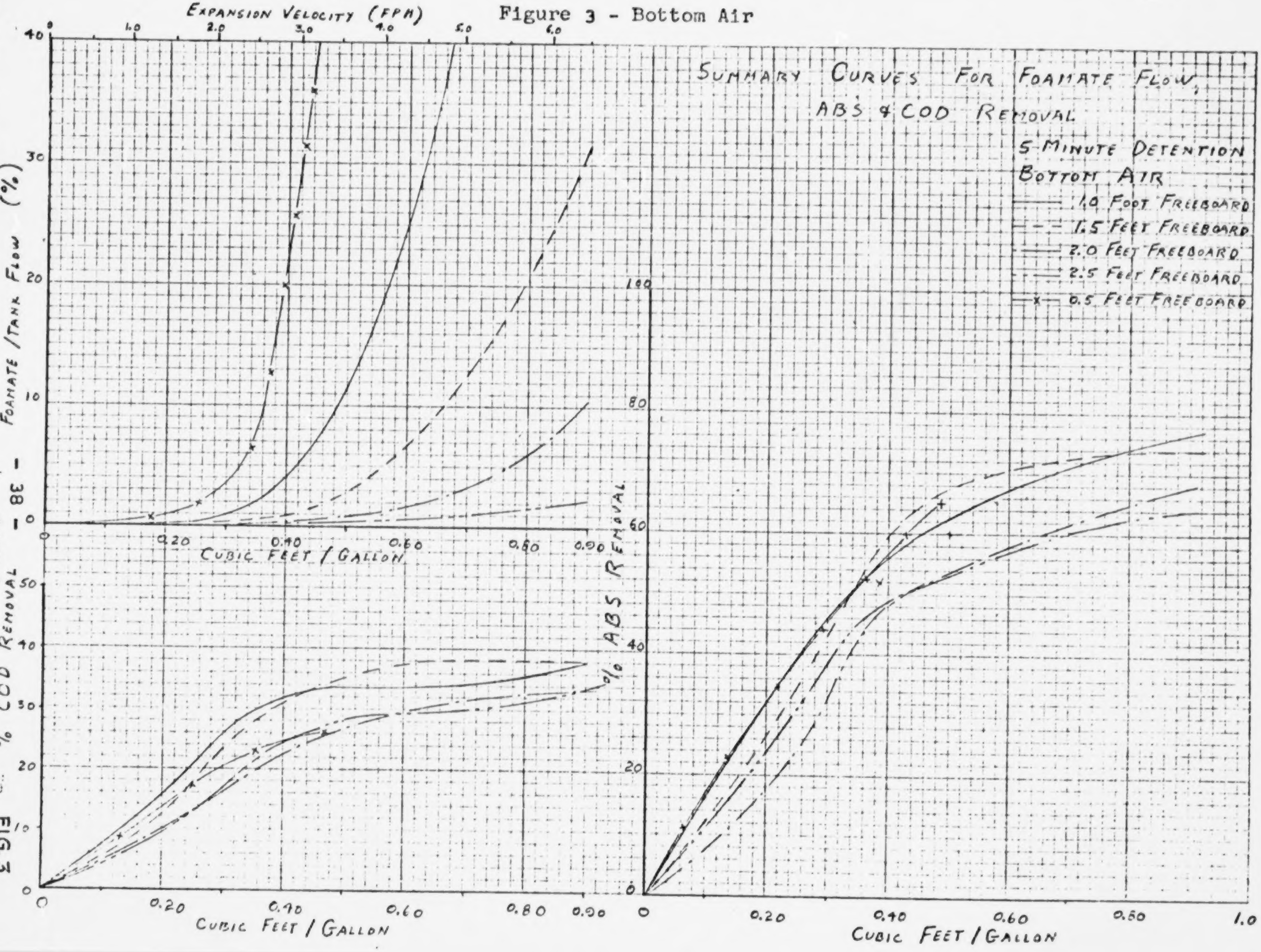

FIG. 3

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Figure 4 - Bottom Air

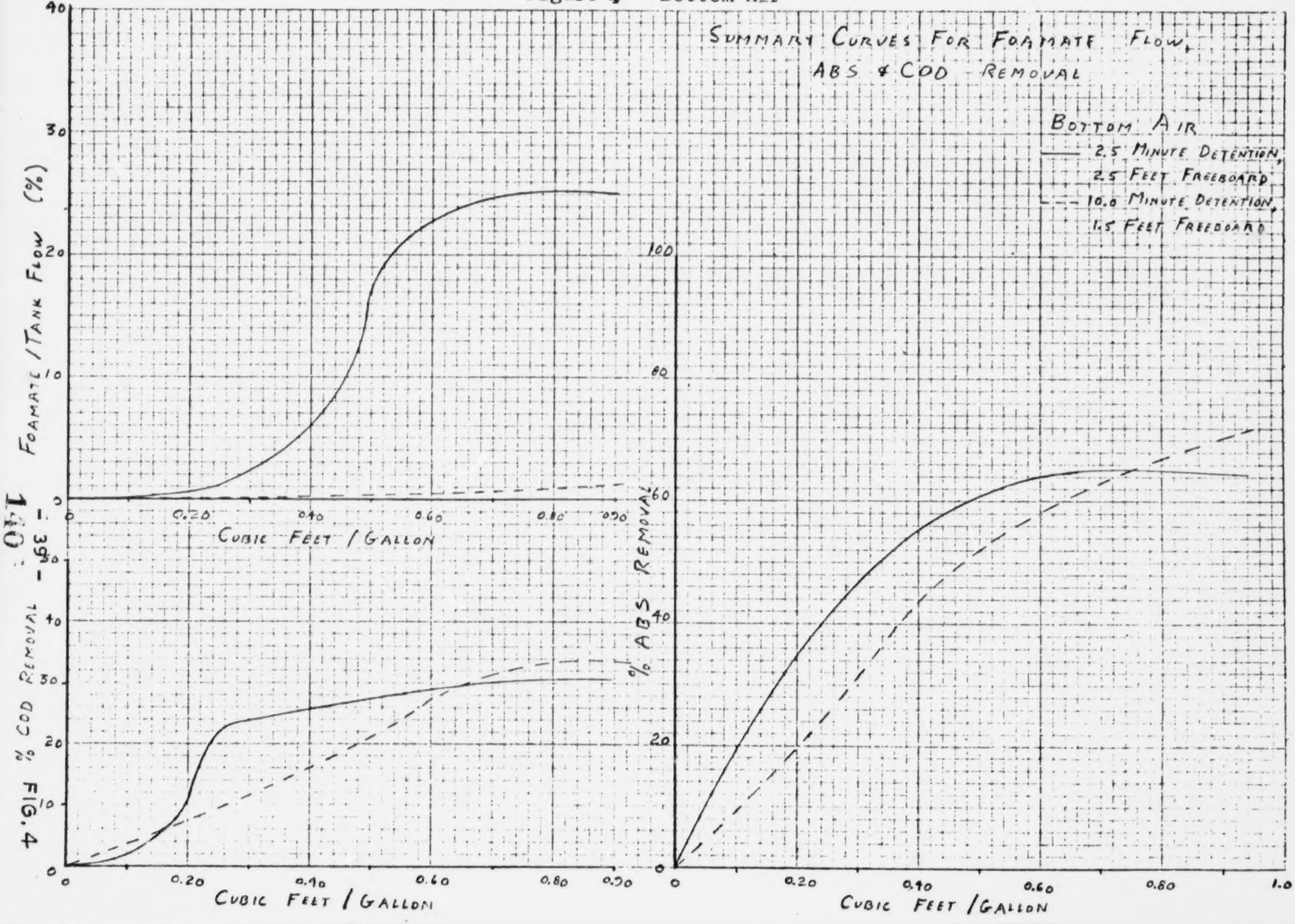

FIG. 4

Figure 5 - Middle Air

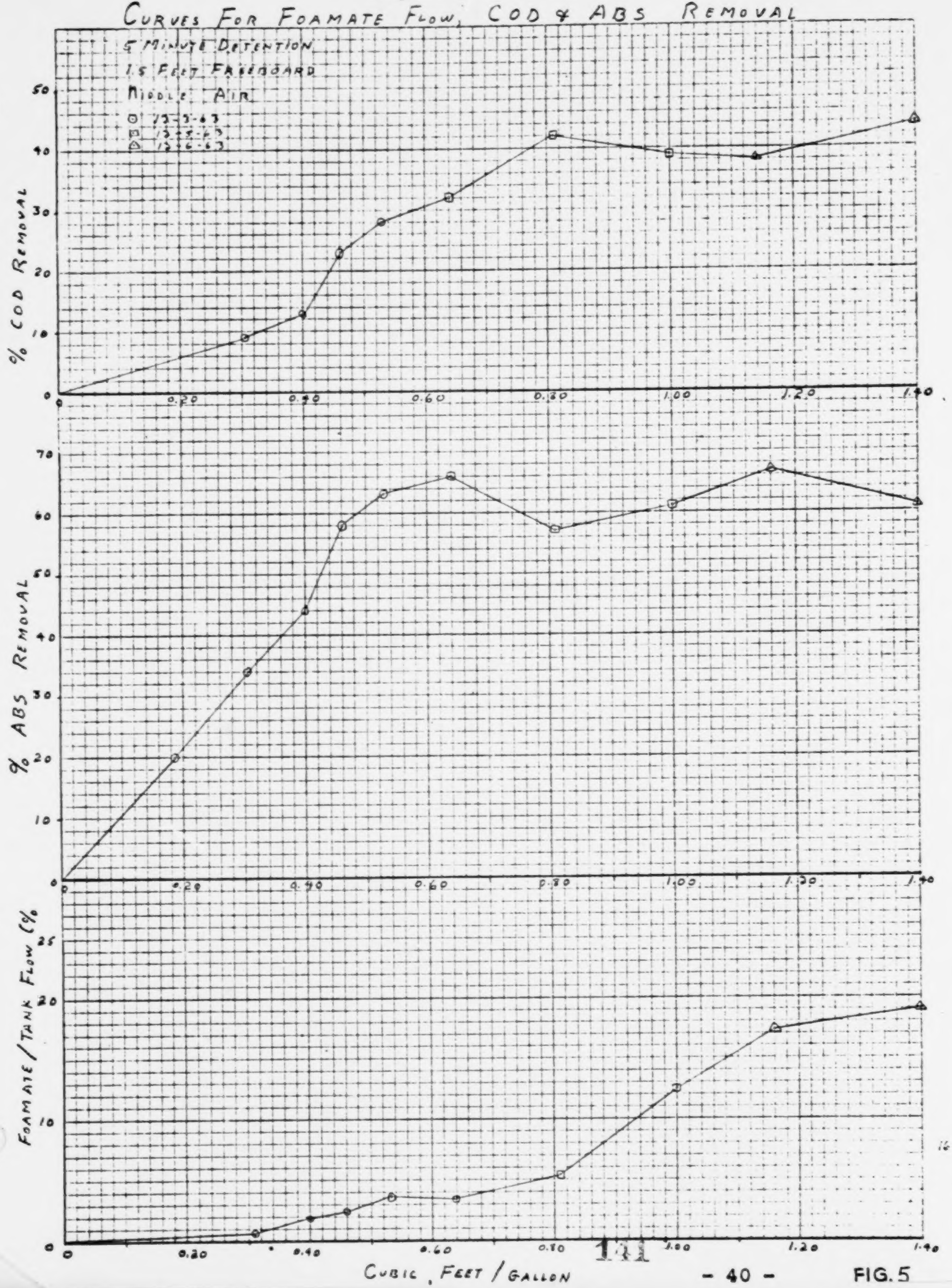

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FIG. 5

Figure 6 - Top Air

CURVES FOR FOAMATE FLOW, COD & ABS REMOVAL

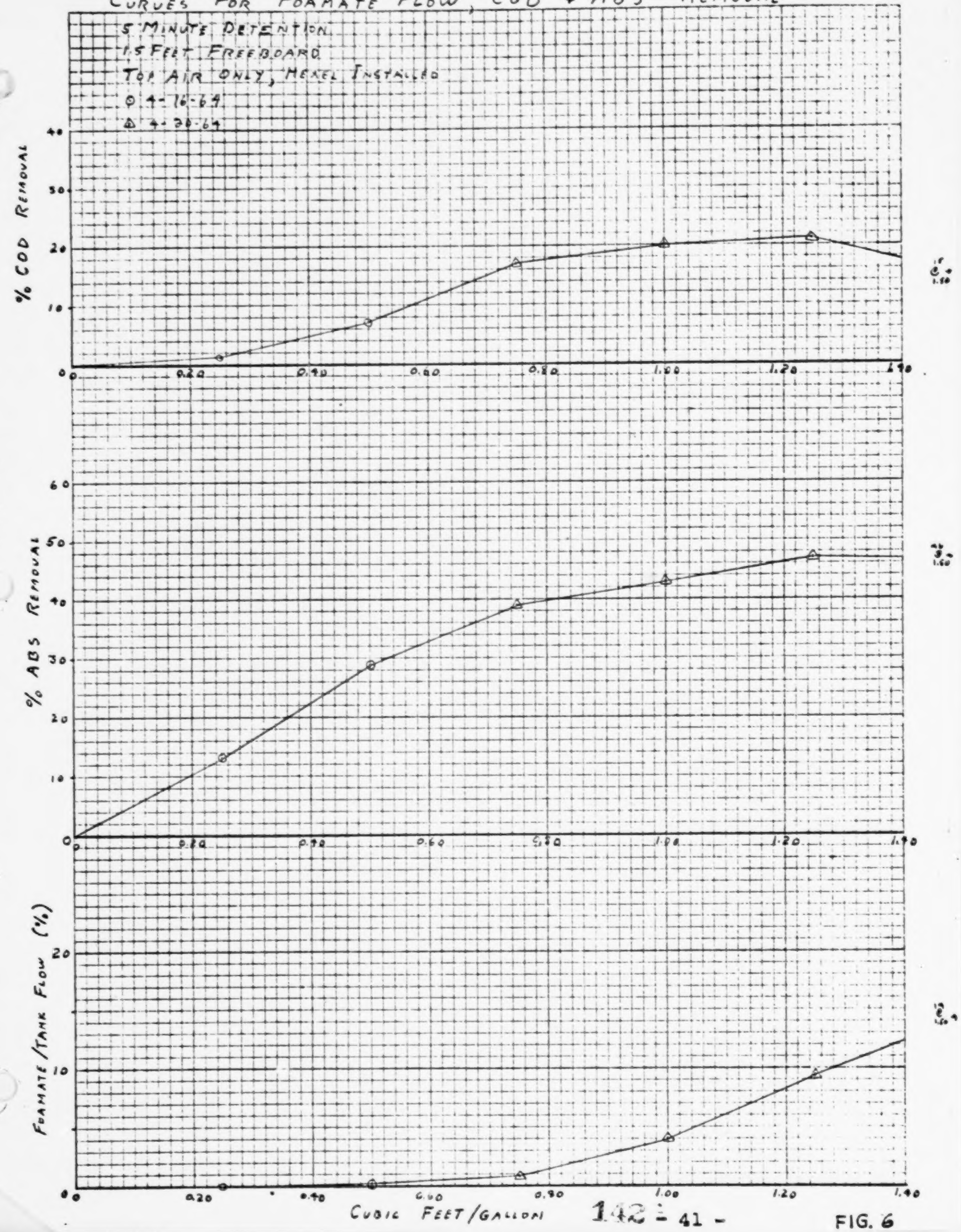

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FIG. 6

Figure 7 - Split Air - (Bottom & Middle)

Curves For Foamate Flow, COD & ABS Removal

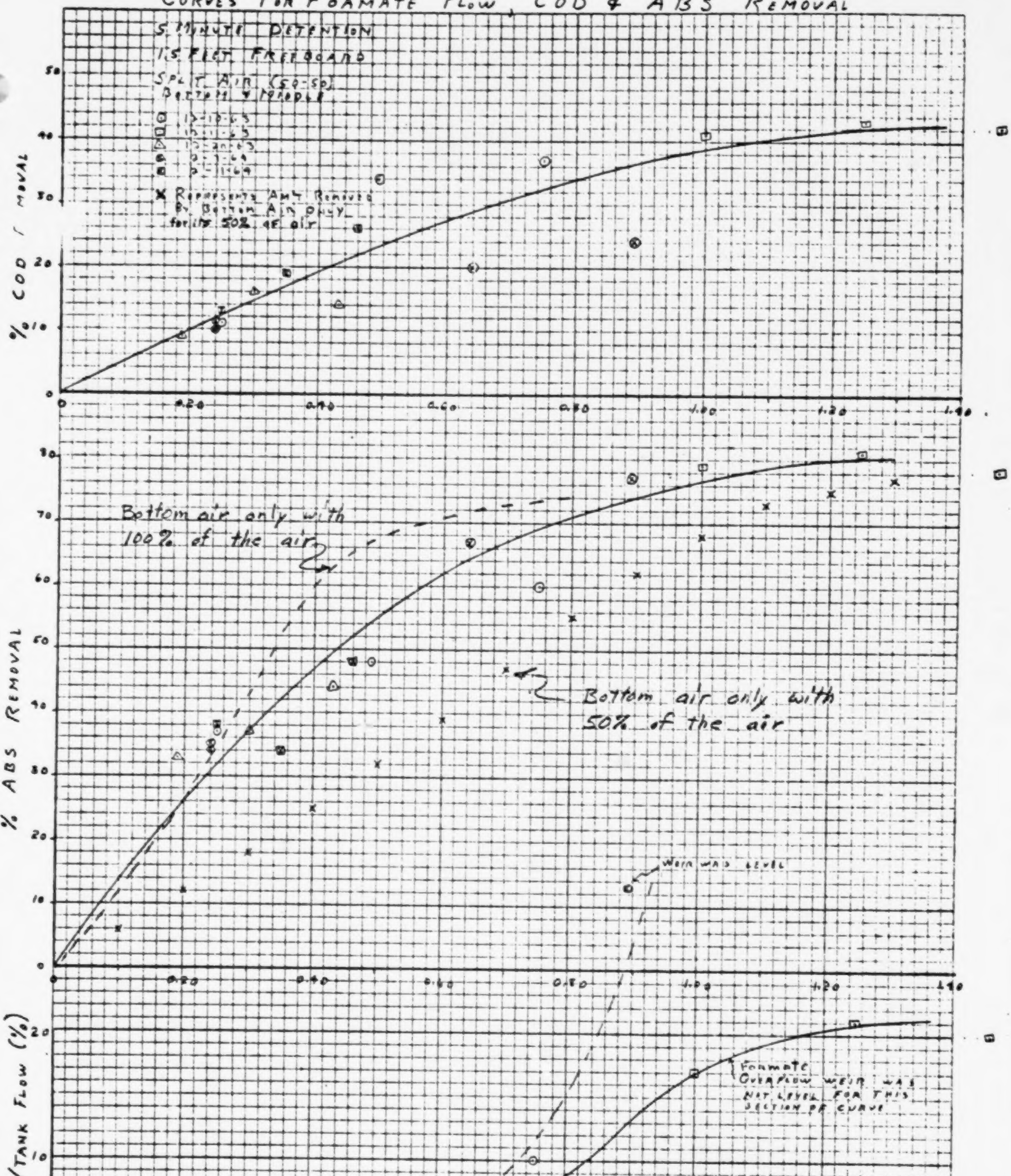

FIG. 7

Figure 8 - Split Air - (Bottom & Top)

CURVES FOR FOAMATE FLOW, COD & ABS REMOVAL

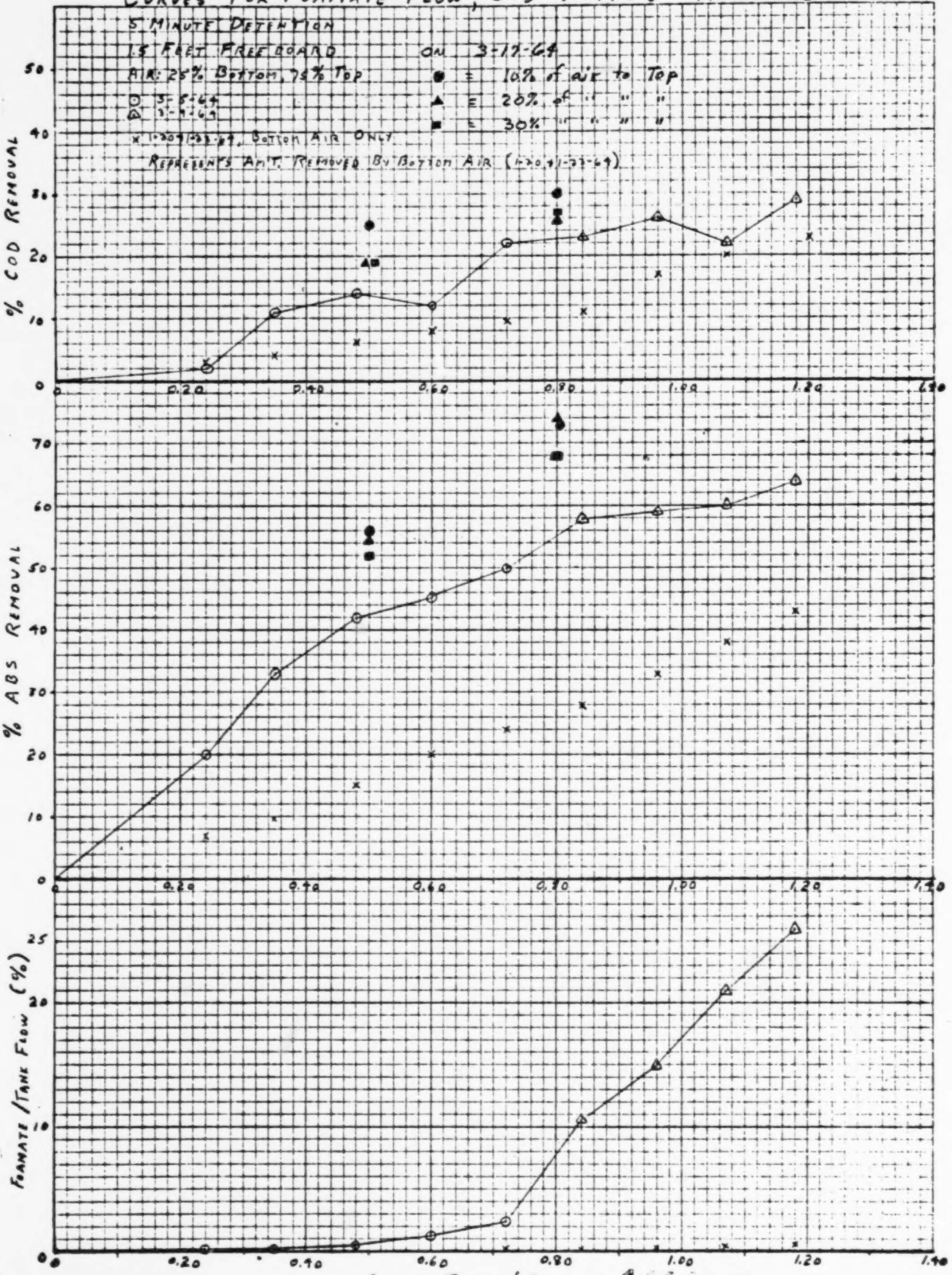

FIG. 8

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Figure 9 - Freeboard

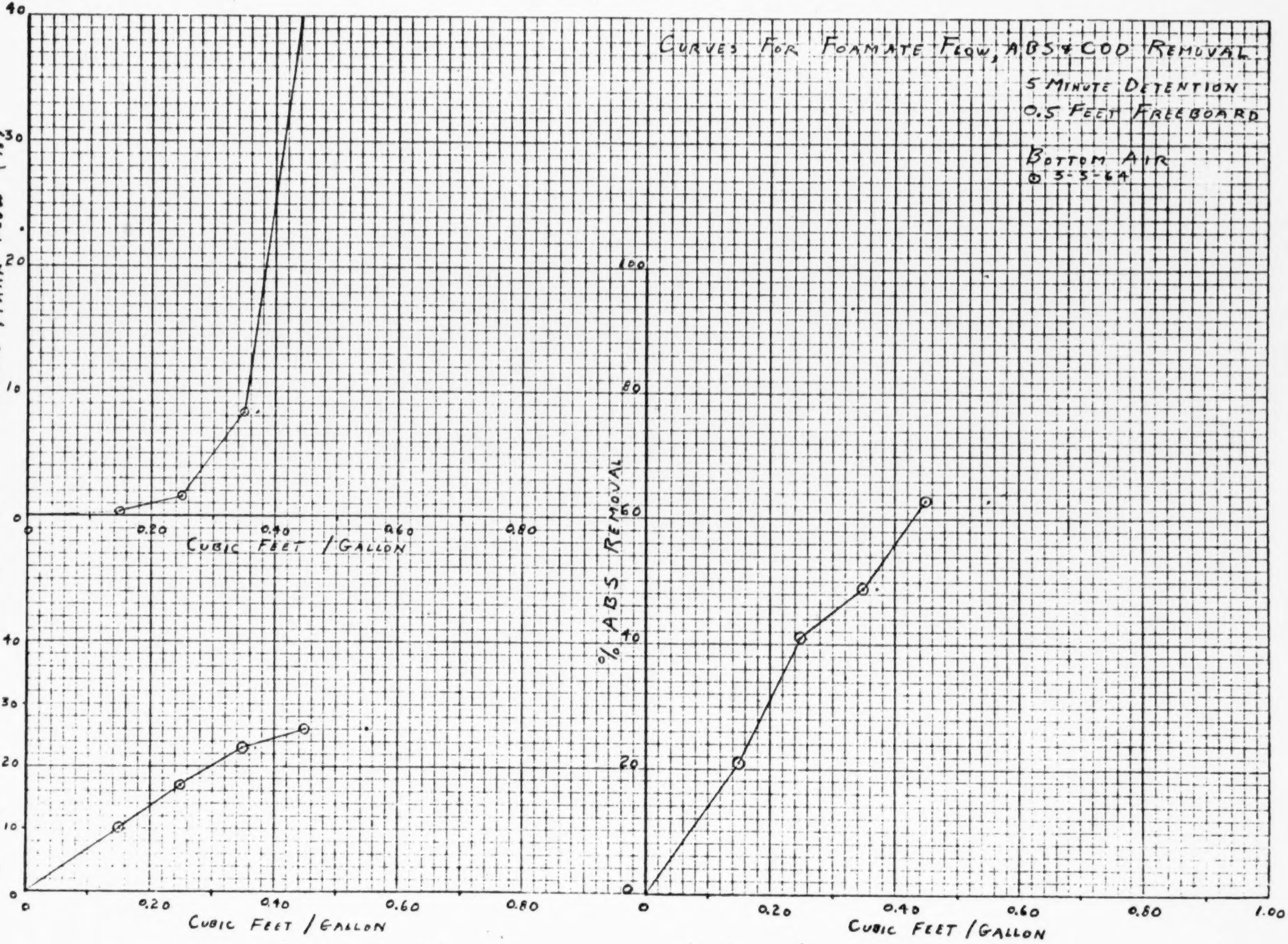

FIG. 9

Figure 10 - Freeboard

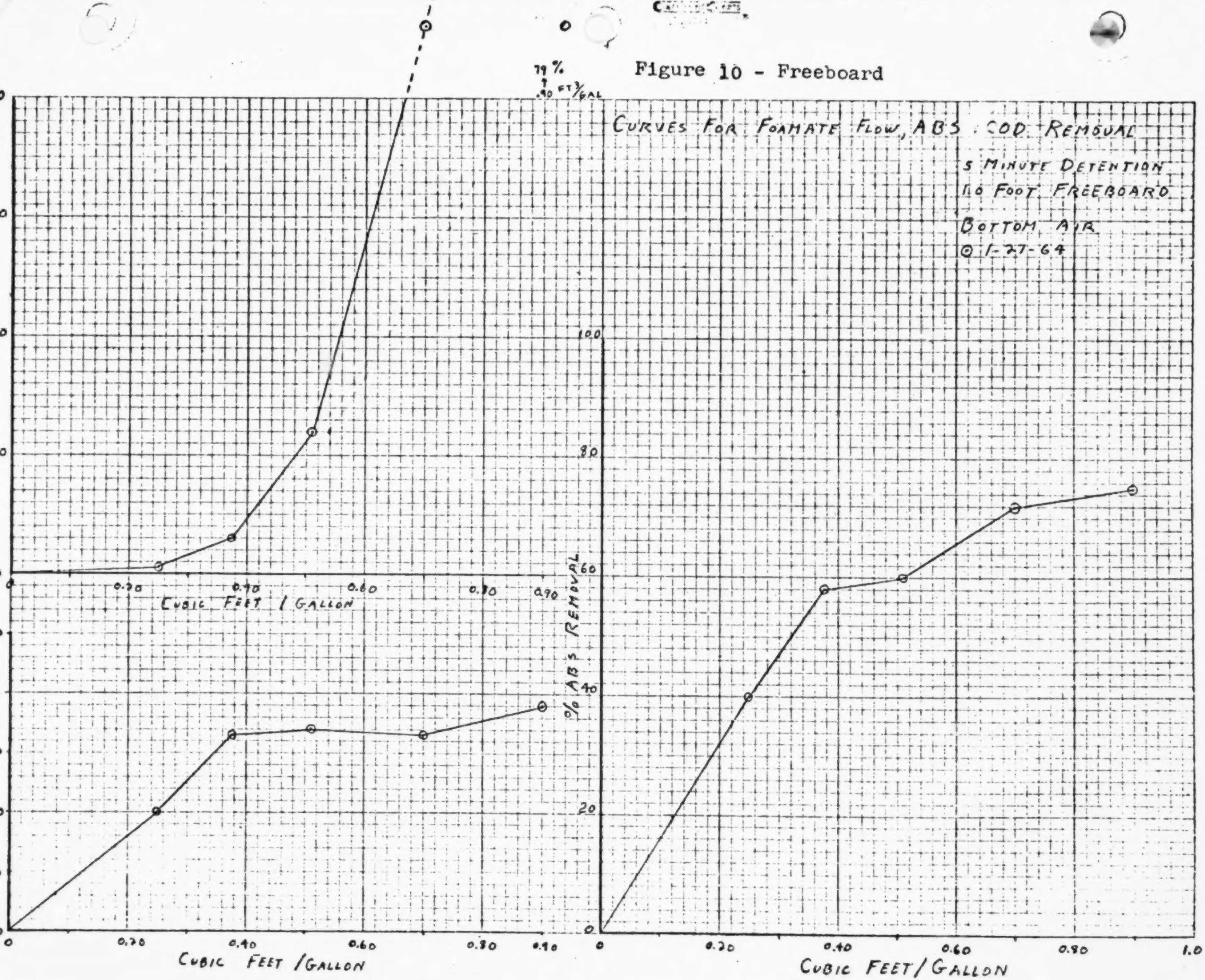

FIG. 10

Figure 11 - Freeboard

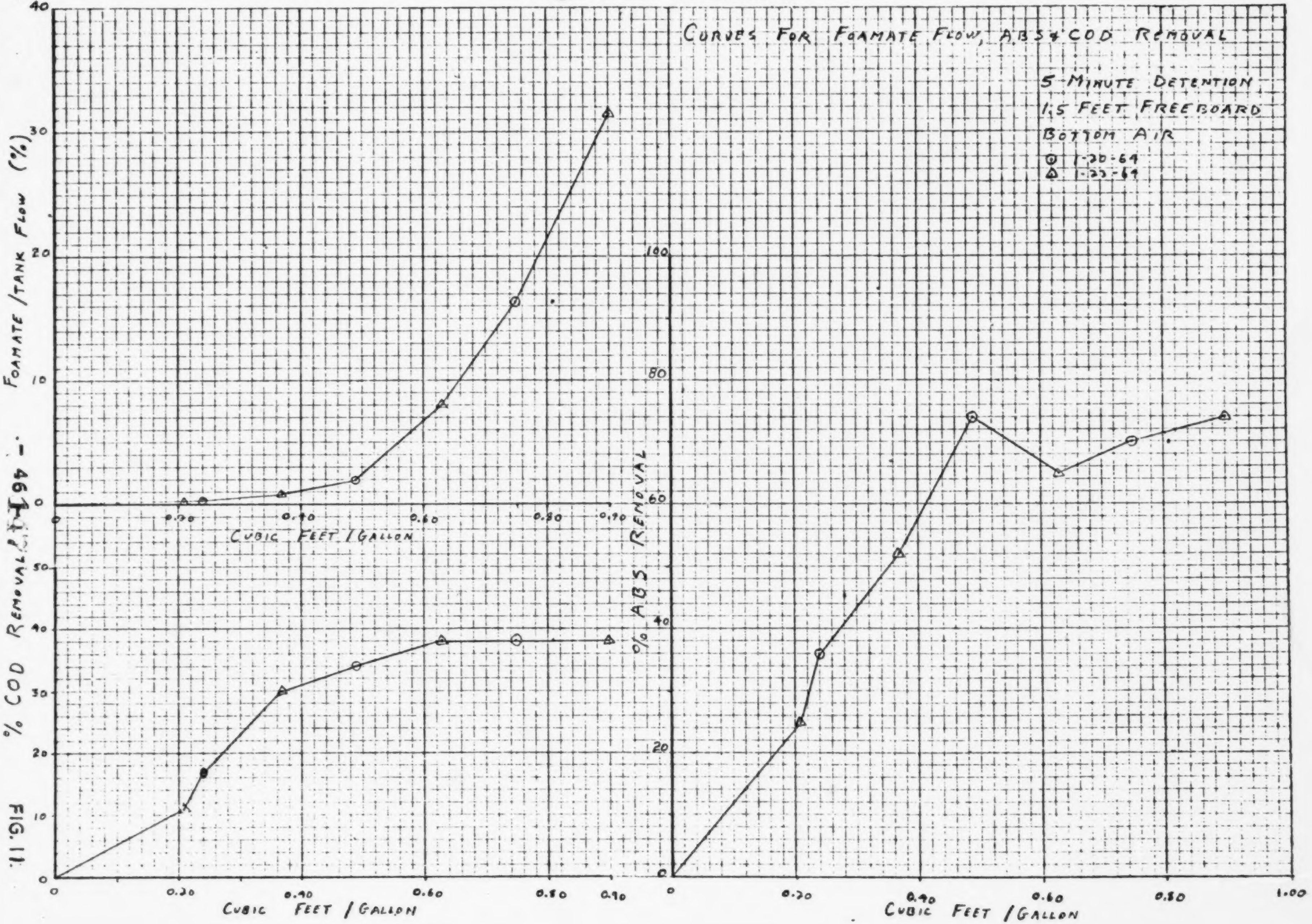

FIG. 11.

Figure: 12 - Freeboard

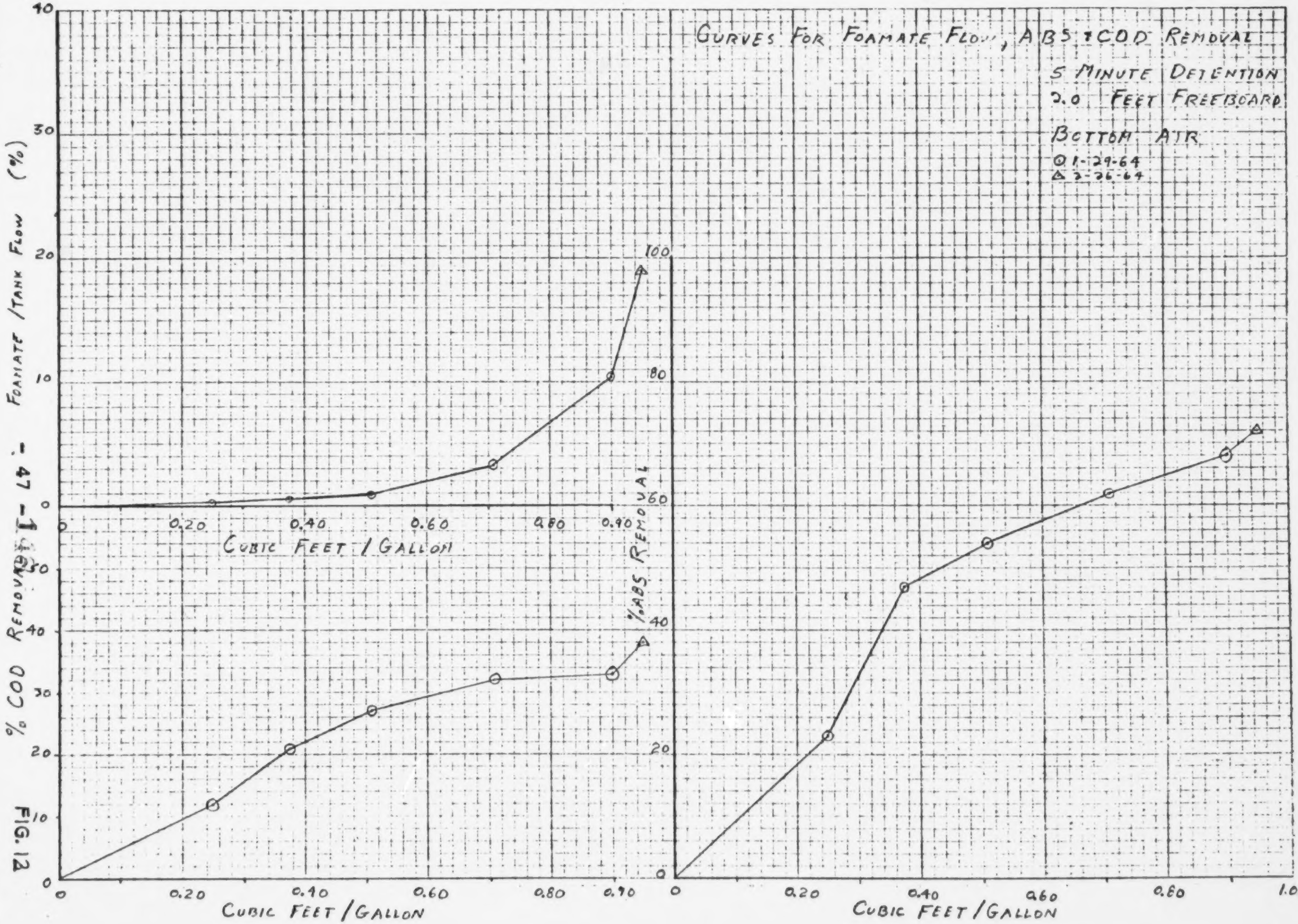

FIG. 12

Figure 13 - Freeboard

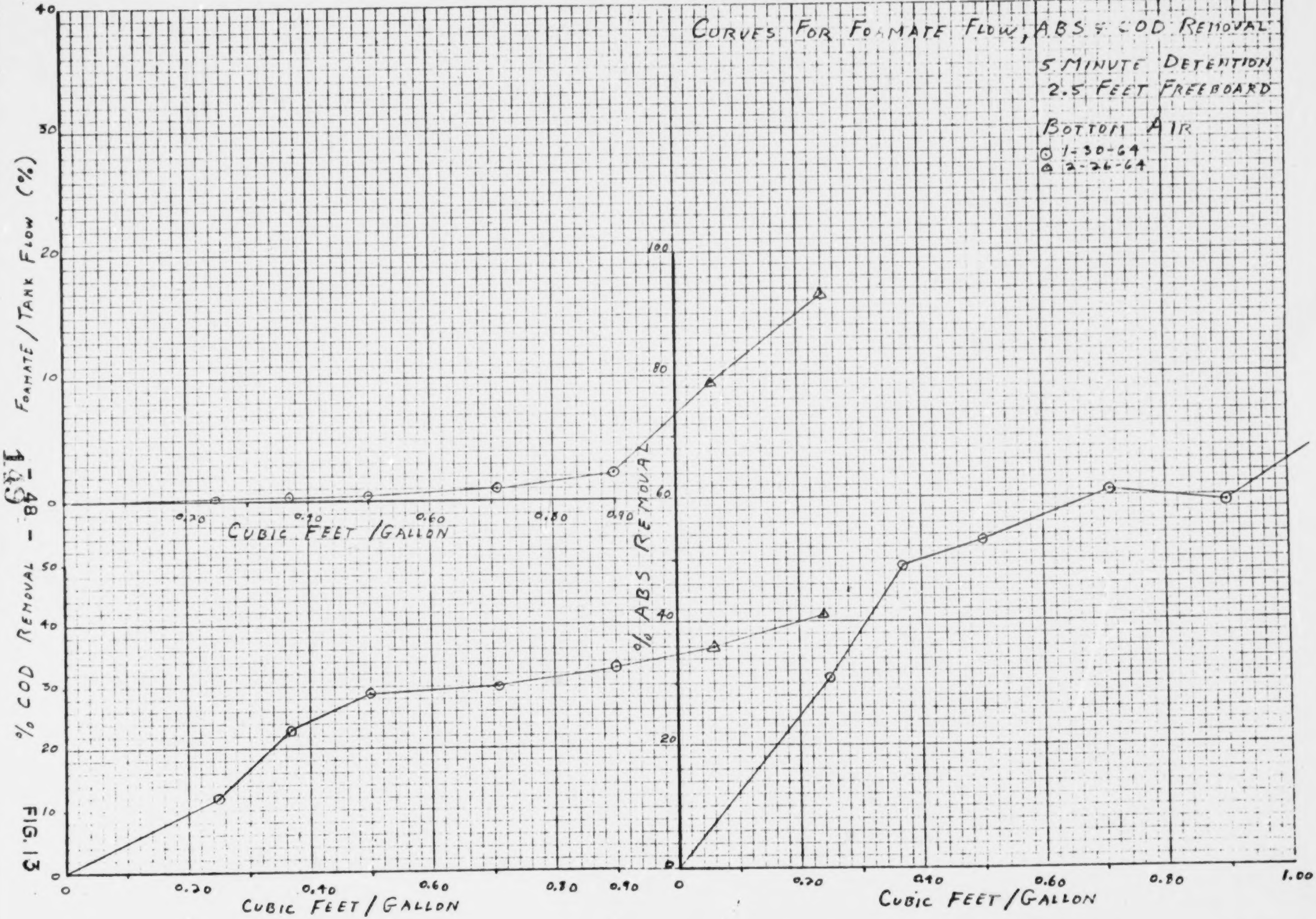

FIG. 13

Figure 14 - Detention Time

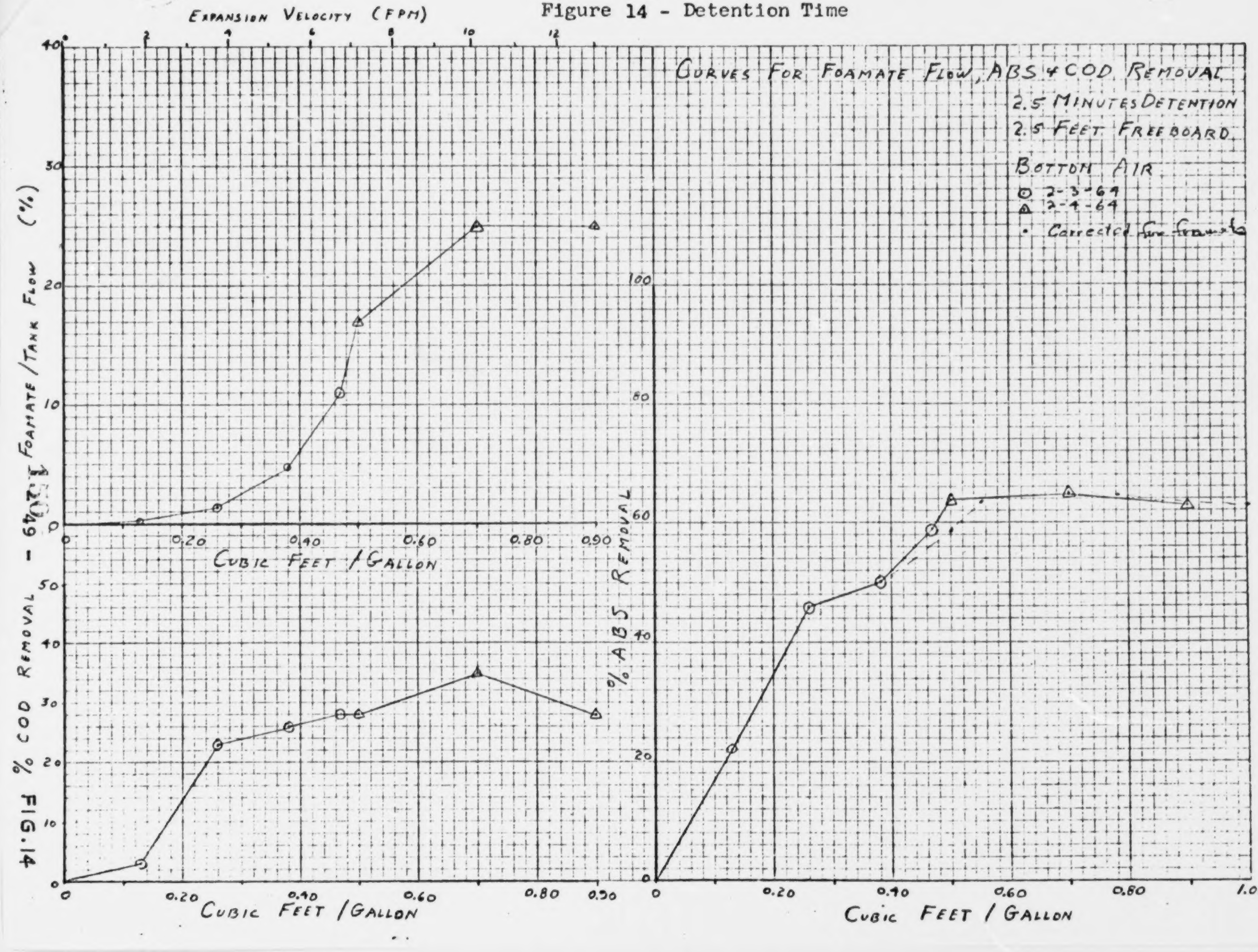

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Figure 15 - Detention Time

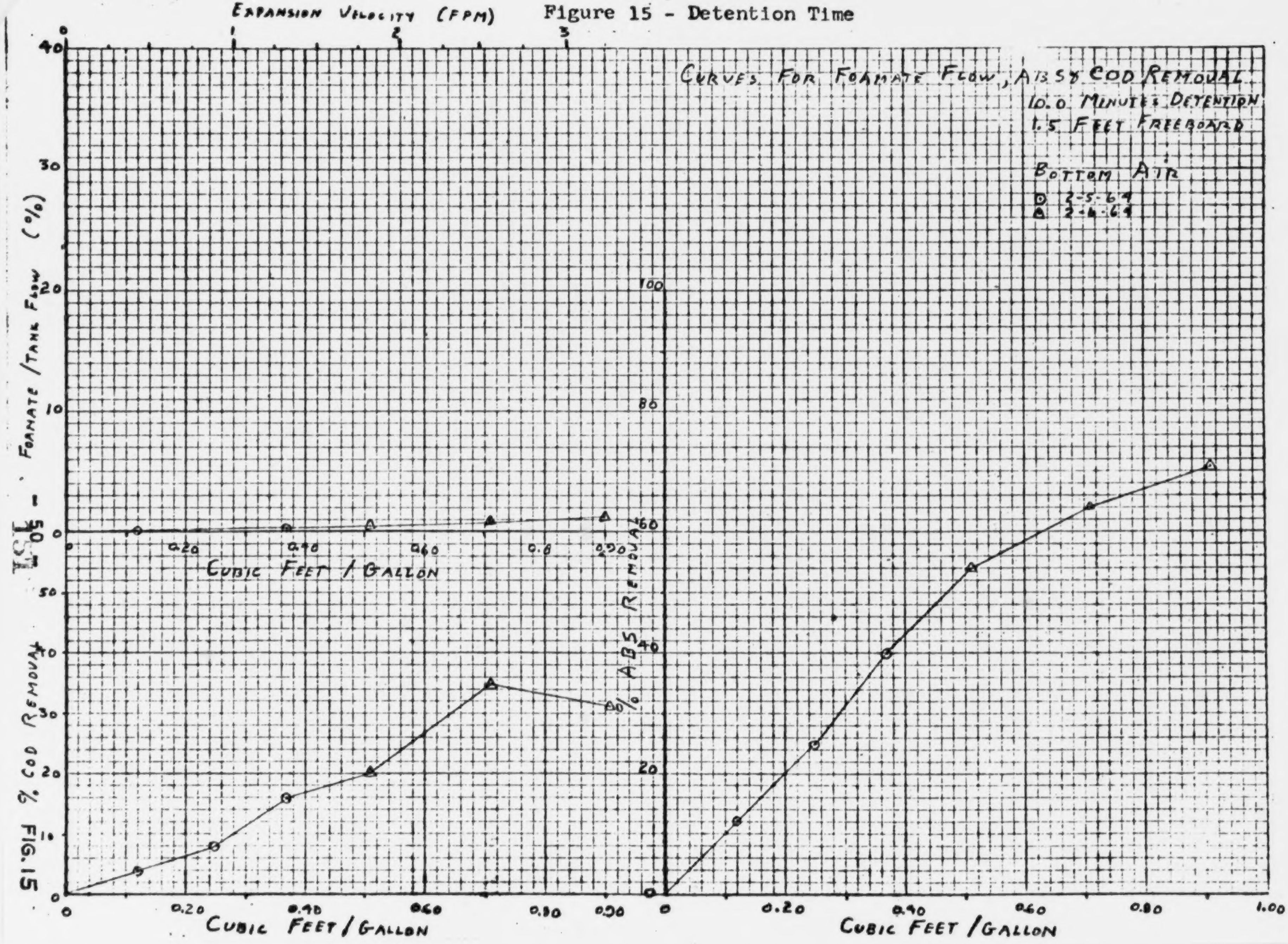


FIG.15

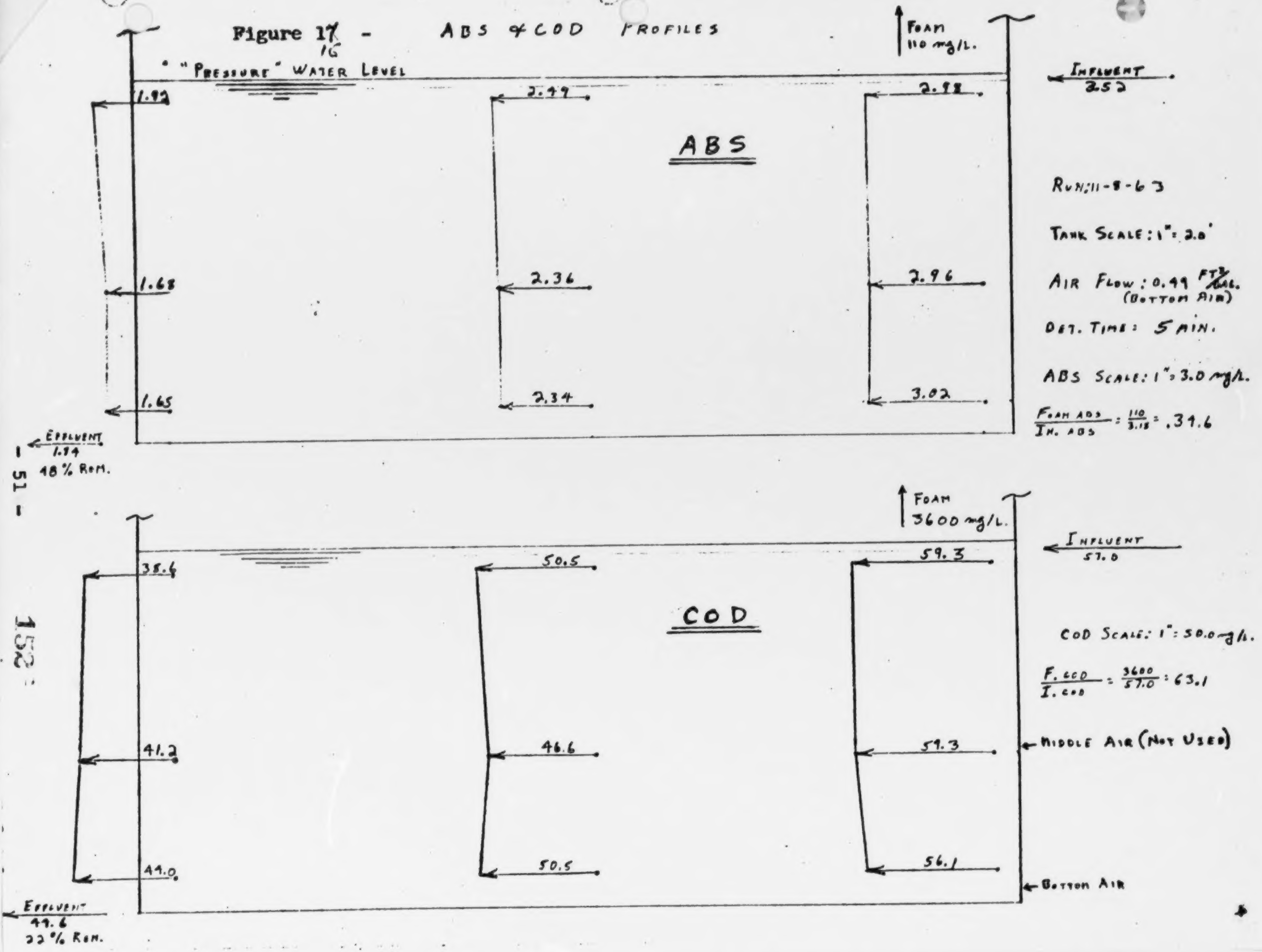
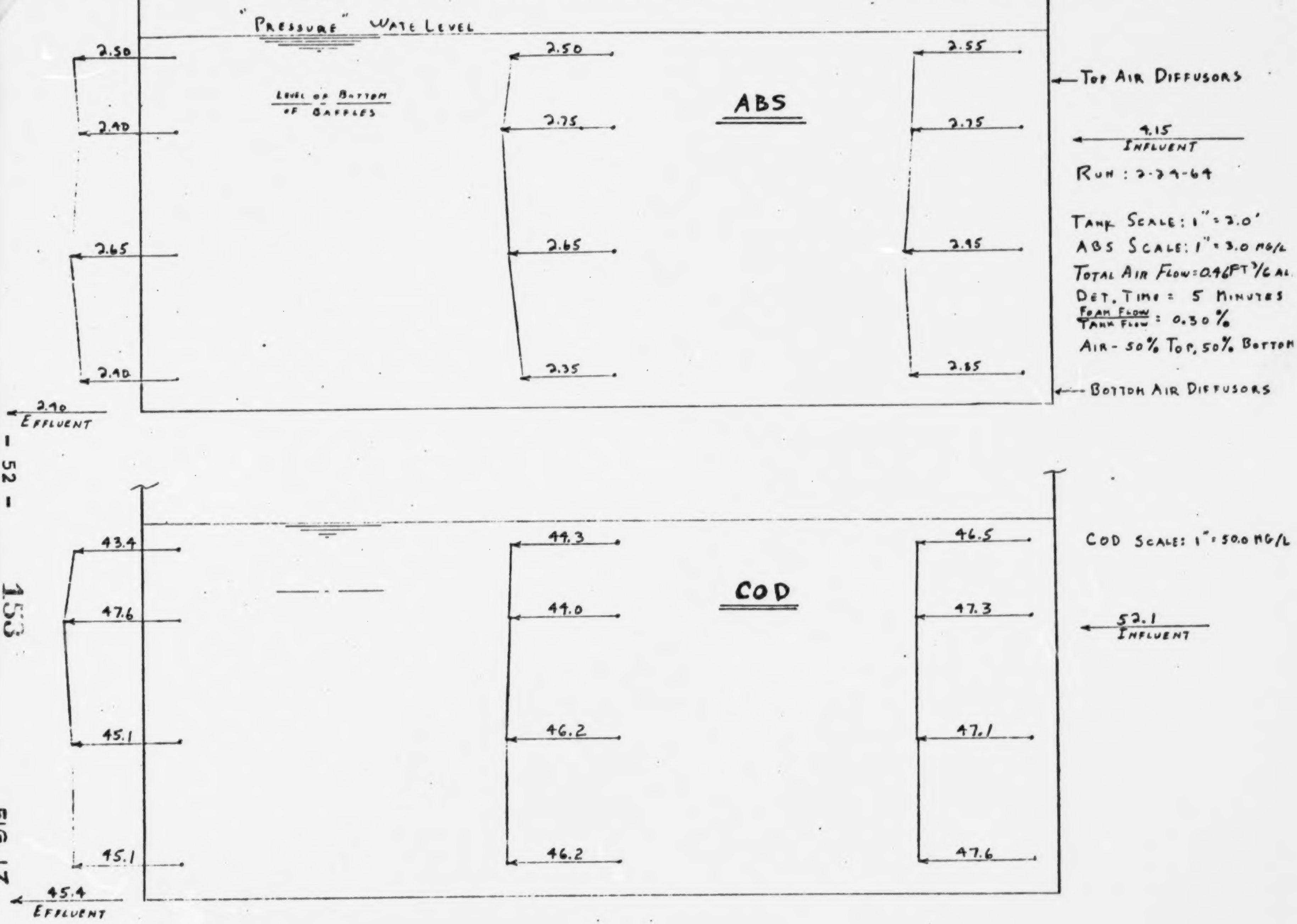


Figure 1~~7~~ 16 - ABS & COD PROFILES

"PRESSURE" WATER LEVEL
FOAM 110 mg/L.
INFLUENT 252
ABS
1.92
1.68
1.65
2.49
2.36
2.34
2.98
2.96
3.02
EFFLUENT 1.34
48 % REM.
RUN: 11-8-63
TANK SCALE: 1" = 2.0'
AIR FLOW: 0.49 FT³/GAL. (BOTTOM AIR)
DET. TIME: 5 MIN.
ABS SCALE: 1" = 3.0 mg/L.
FOAM ABS / IN. ABS = 110 / 3.18 = 34.6
FOAM 3600 mg/L.
INFLUENT 57.0
COD
35.6
41.2
44.0
50.5
46.6
50.5
59.3
59.3
56.1
EFFLUENT 44.6
22 % REM.
COD SCALE: 1" = 50.0 mg/L.
F. COD / I. COD = 3600 / 57.0 = 63.1
MIDDLE AIR (NOT USED)
BOTTOM AIR

Figure ~~18~~ 17 - ABS & COD Profiles - Air 50% to Top - 50% to Botto


"Pressure" Wate Level
Level of Bottom of Baffles
ABS
2.50
2.40
2.65
2.40
2.40 Effluent
2.50
2.75
2.65
2.35
2.55
2.75
2.95
2.85
Top Air Diffusors
4.15 Influent
Bottom Air Diffusors
COD
43.4
47.6
45.1
45.1
45.4 Effluent
44.3
44.0
46.2
46.2
46.5
47.3
47.1
47.6
52.1 Influent

Run : 2-29-64

Tank Scale: 1" = 2.0'

ABS Scale: 1" = 3.0 MG/L

Total Air Flow = 0.46 FT³/Gal.

Det. Time = 5 Minutes

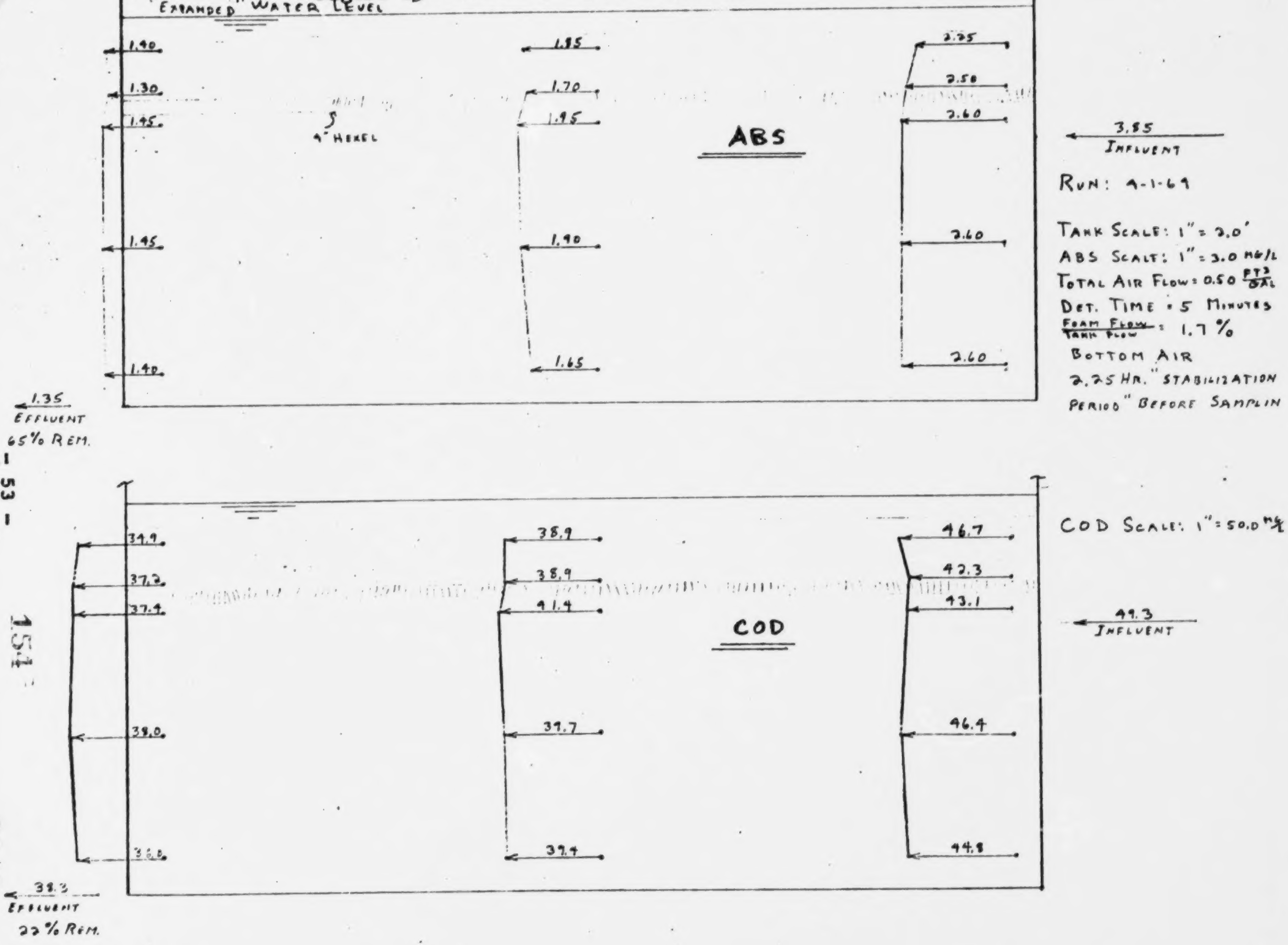
$\frac{\text{Foam Flow}}{\text{Tank Flow}}$ = 0.30 %

Air - 50% Top, 50% Bottom

COD Scale: 1" = 50.0 MG/L

18

Figure ~~19~~ - ABS + COD PROFILES


"EXPANDED" WATER LEVEL
4" HEXEL
ABS
1.40
1.30
1.45
1.45
1.40
1.35
EFFLUENT
65% REM.
1.85
1.70
1.95
1.90
1.65
2.25
2.50
2.60
2.60
2.60
3.85
INFLUENT
COD
34.9
37.2
37.4
39.0
36.8
38.3
EFFLUENT
22% REM.
38.9
38.9
41.4
39.7
39.4
46.7
42.3
43.1
46.4
44.8
49.3
INFLUENT

RUN: 4-1-69

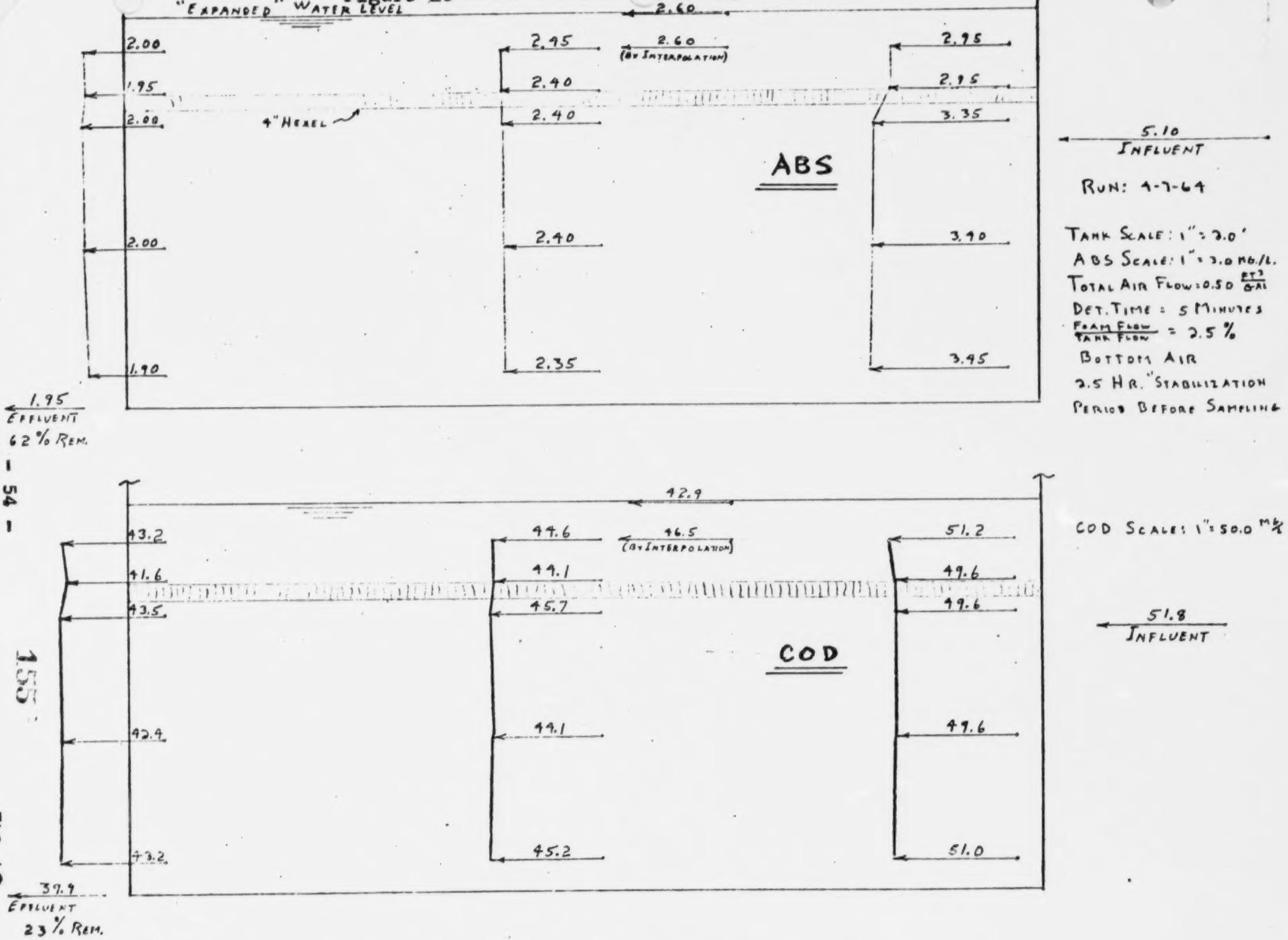
TANK SCALE: 1" = 2.0'
ABS SCALE: 1" = 3.0 MG/L
TOTAL AIR FLOW = 0.50 $\frac{FT^3}{GAL}$
DET. TIME = 5 MINUTES
$\frac{\text{FOAM FLOW}}{\text{TANK FLOW}}$ = 1.7 %
BOTTOM AIR
2.25 HR. "STABILIZATION PERIOD" BEFORE SAMPLIN

COD SCALE: 1" = 50.0 MG/L

FIG. 18

17

Figure 20 ABS & COD ROFILES


"EXPANDED" WATER LEVEL
2.60
2.00
1.95
2.00
2.00
1.90
2.45
2.40
2.40
2.40
2.35
2.60
(BY INTERPOLATION)
4" HEXEL
2.95
2.95
3.35
3.90
3.95
ABS
5.10
INFLUENT
1.95
EFFLUENT
62 % REM.
42.9
43.2
41.6
43.5
42.4
43.2
44.6
44.1
45.7
44.1
45.2
46.5
(BY INTERPOLATION)
51.2
49.6
49.6
49.6
51.0
COD
51.8
INFLUENT
39.9
EFFLUENT
23 % REM.

RUN: 4-7-64

TANK SCALE: 1" = 2.0'
ABS SCALE: 1" = 2.0 MG./L.
TOTAL AIR FLOW: 0.50 $\frac{FT^3}{GAL}$
DET. TIME = 5 MINUTES
$\frac{\text{FOAM FLOW}}{\text{TANK FLOW}}$ = 2.5 %
BOTTOM AIR
2.5 HR. "STABILIZATION PERIOD BEFORE SAMPLING

COD SCALE: 1" = 50.0 MG/L

FIG. 19

Figure 20 - Foamate Flow

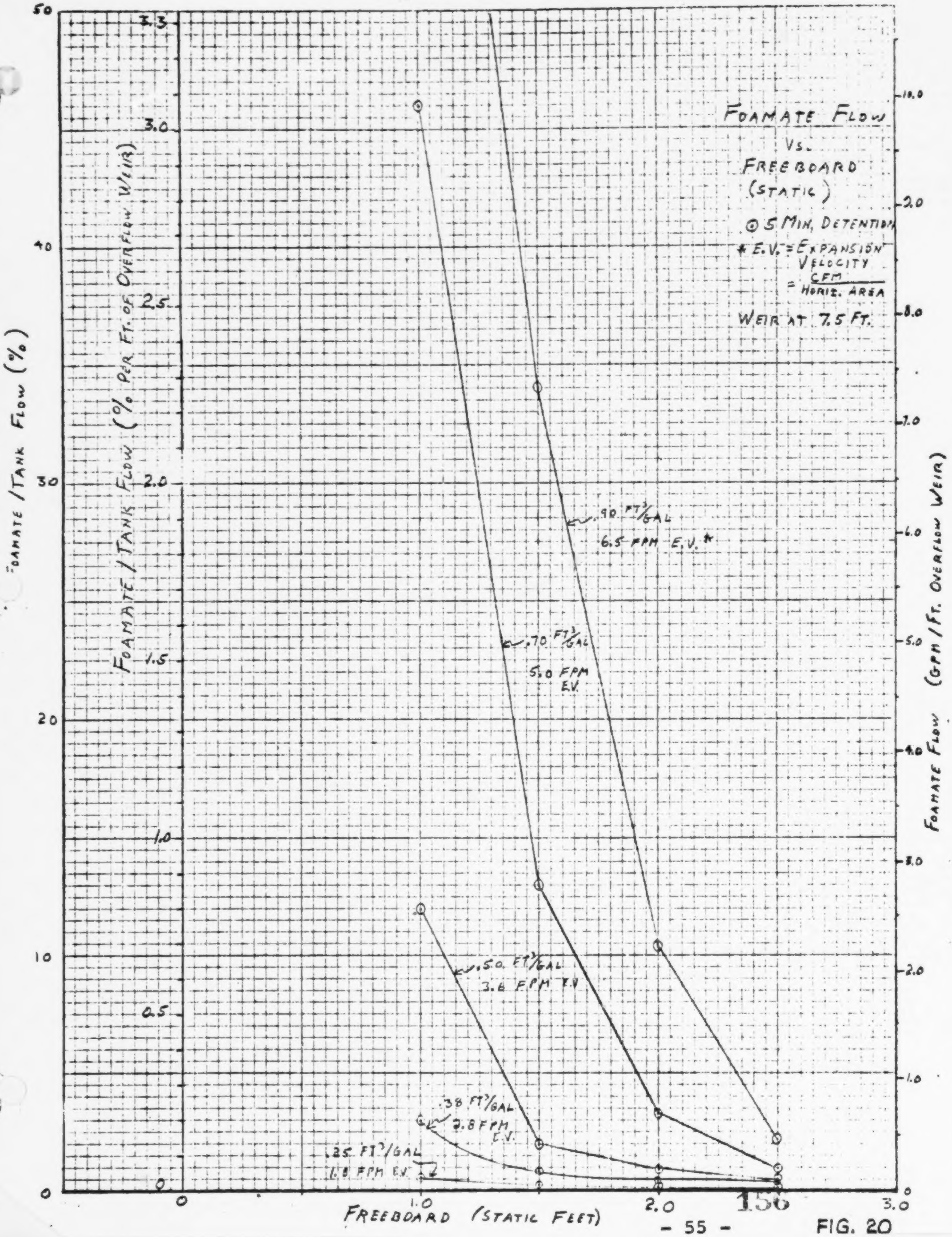


FIG. 20

CURVES FOR FOAMATE FLOW, COD & ABS REMOVAL

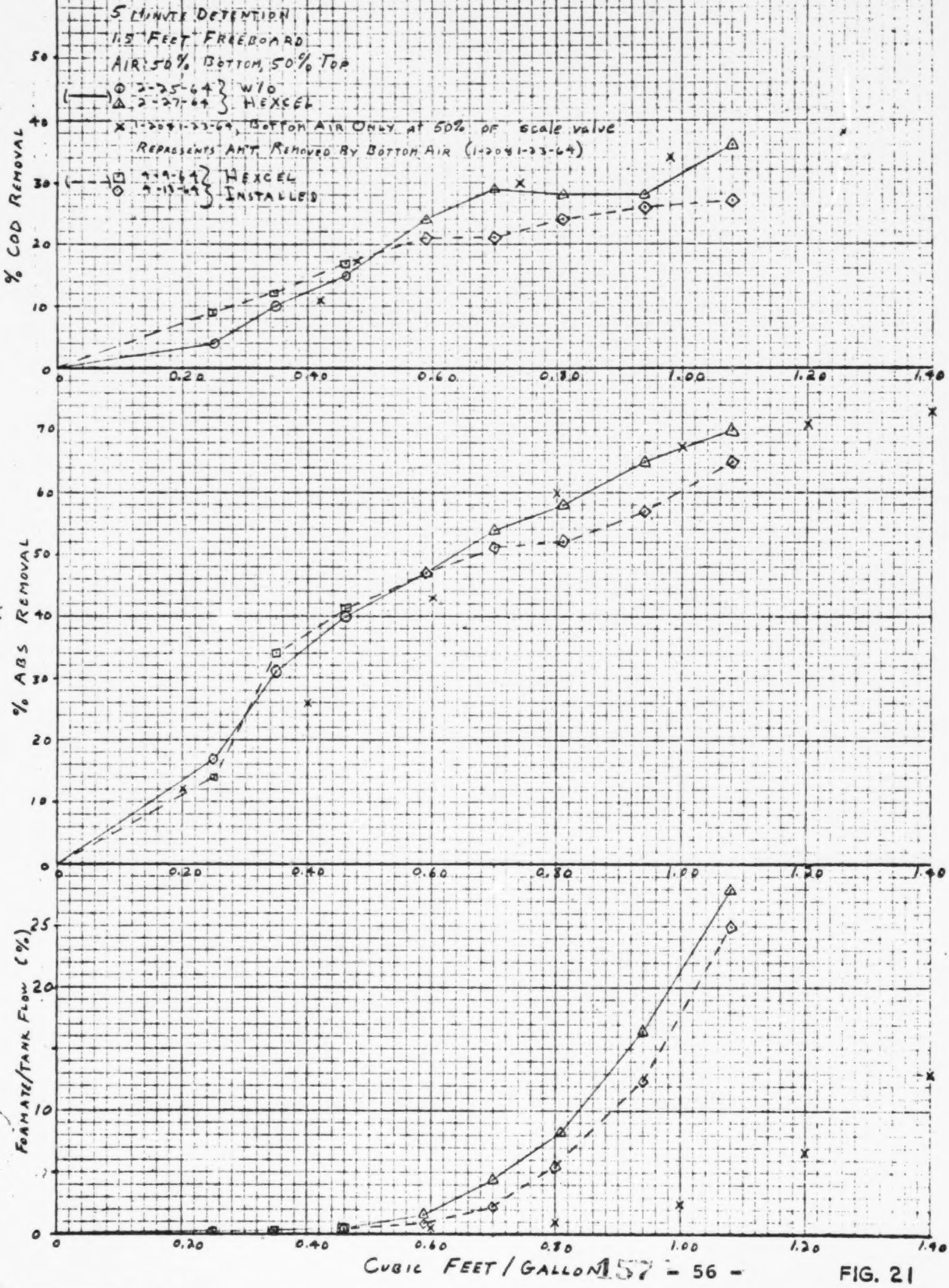


FIG. 21

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Figure 22 - Hexcel

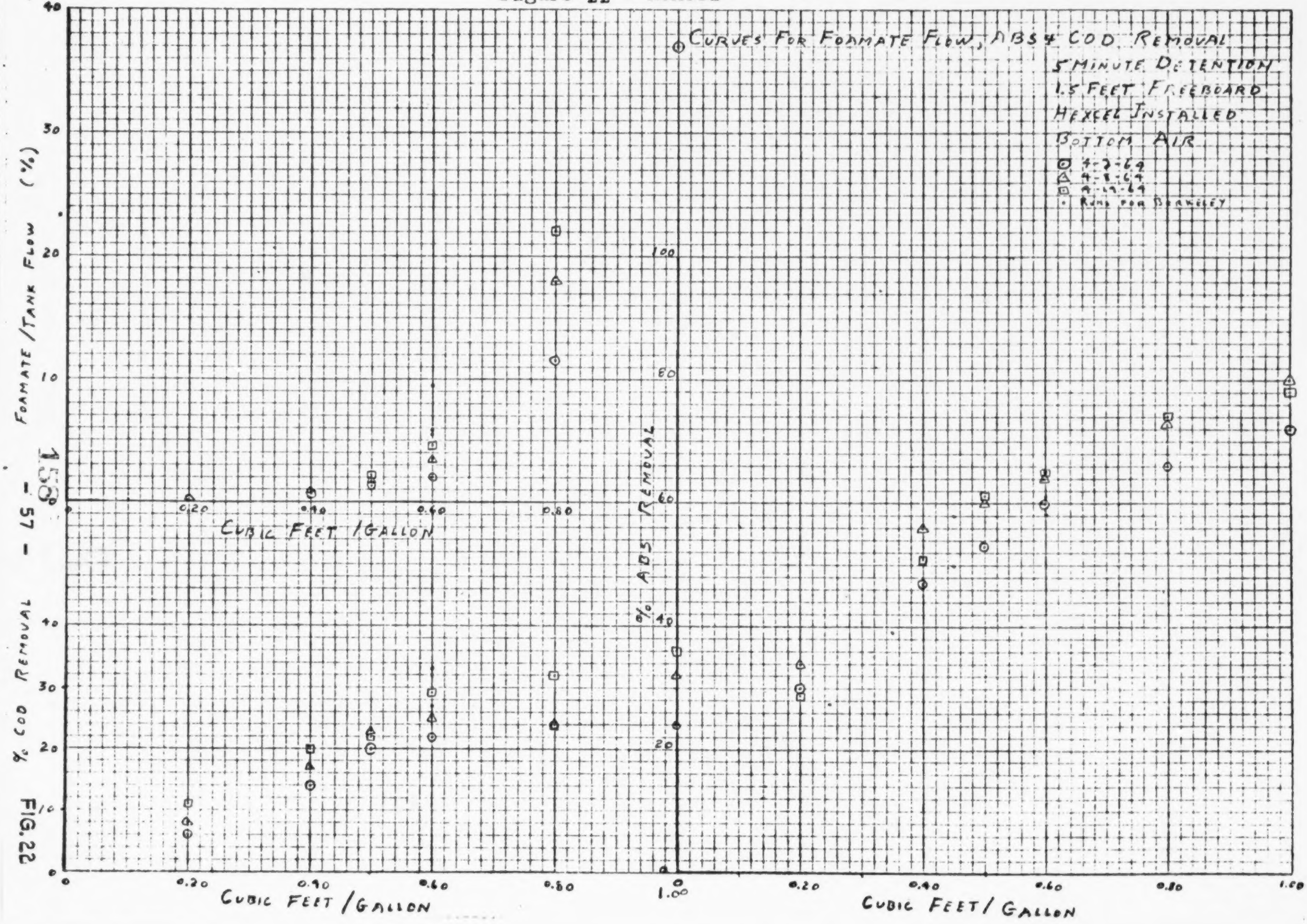

FIG. 22

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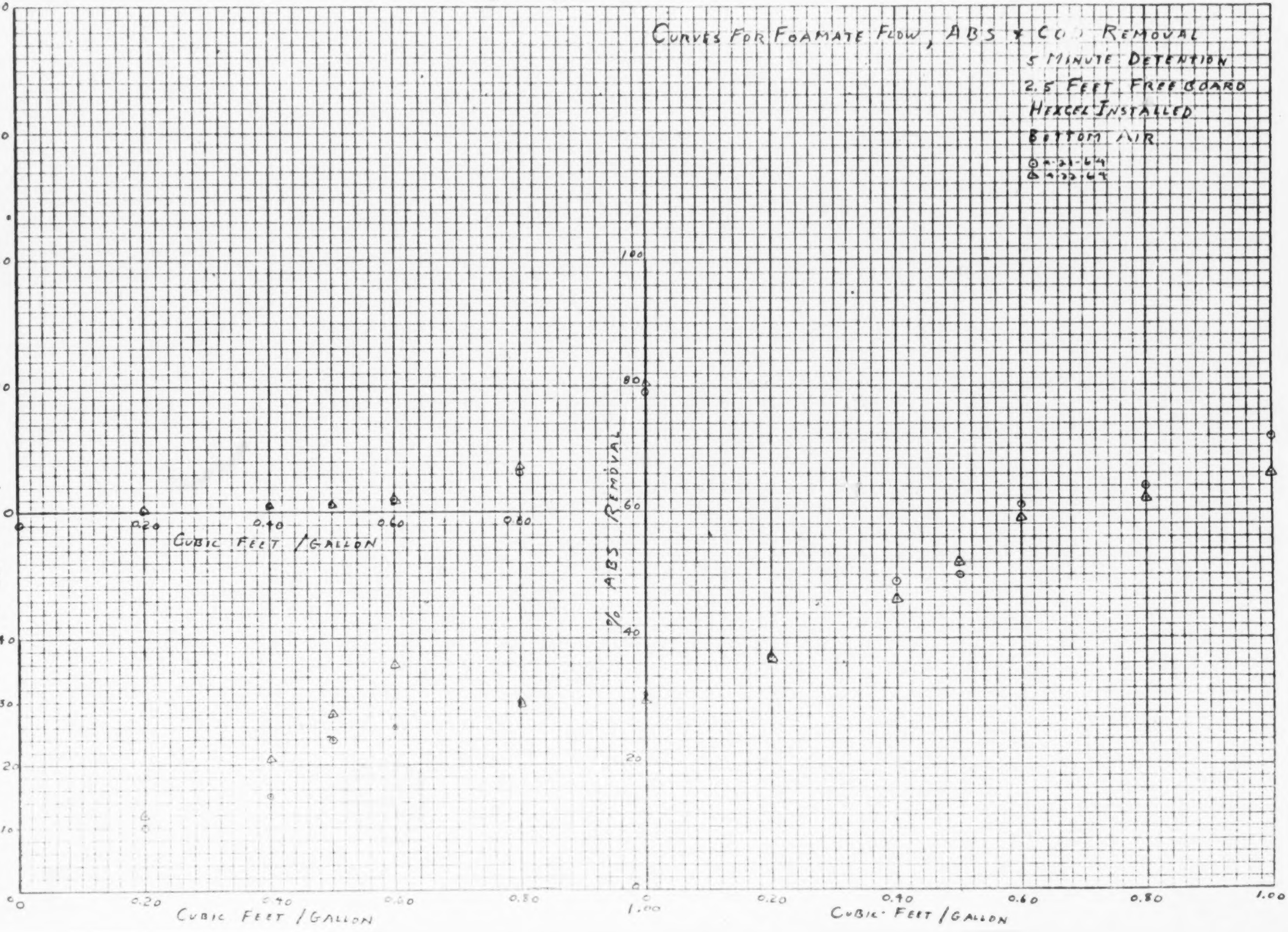
Figure 23 - Hexcel



% COD FIG.23

Figure 24

DYE STUDY - FOAM TANK - ZERO AIR

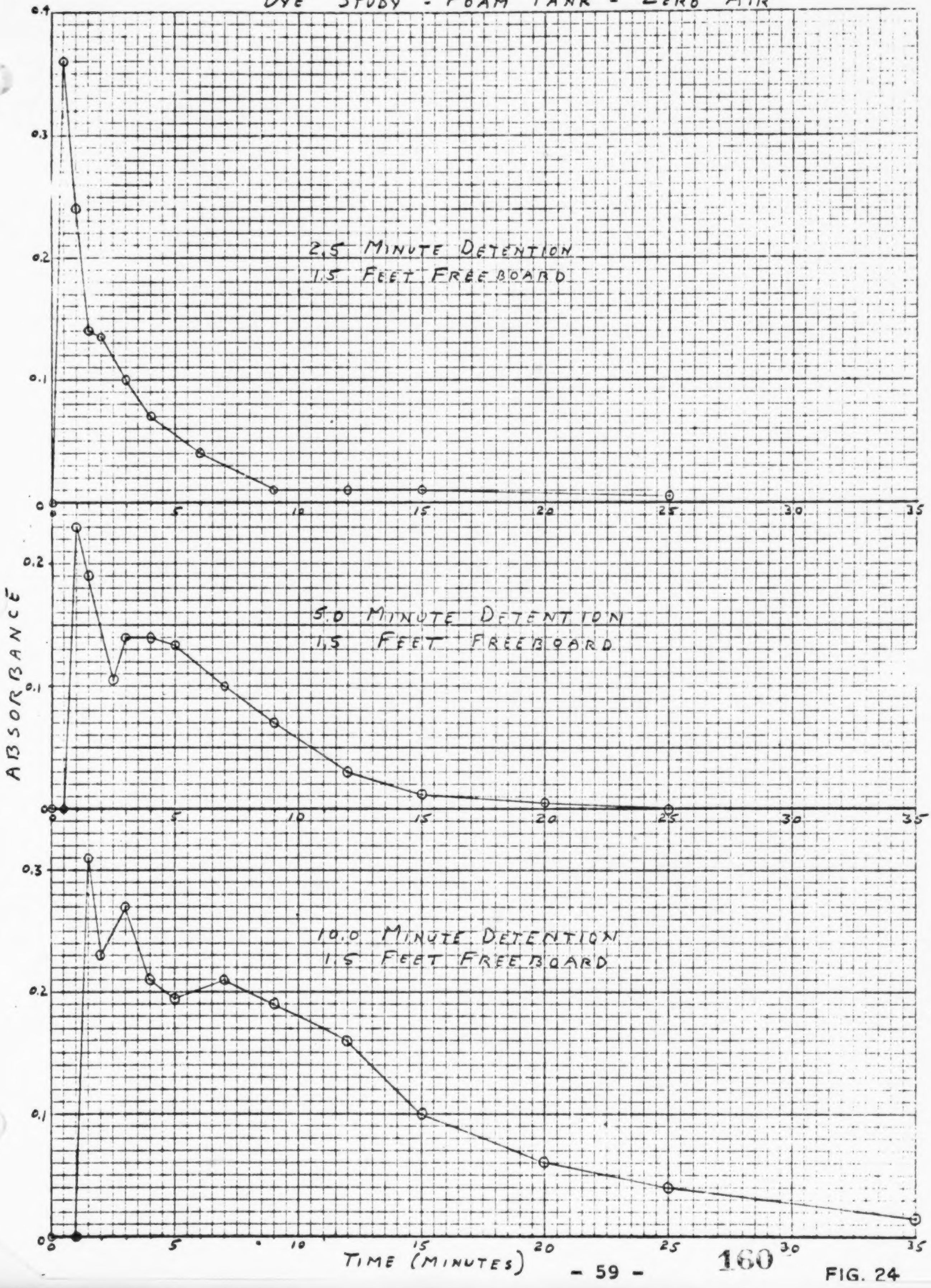
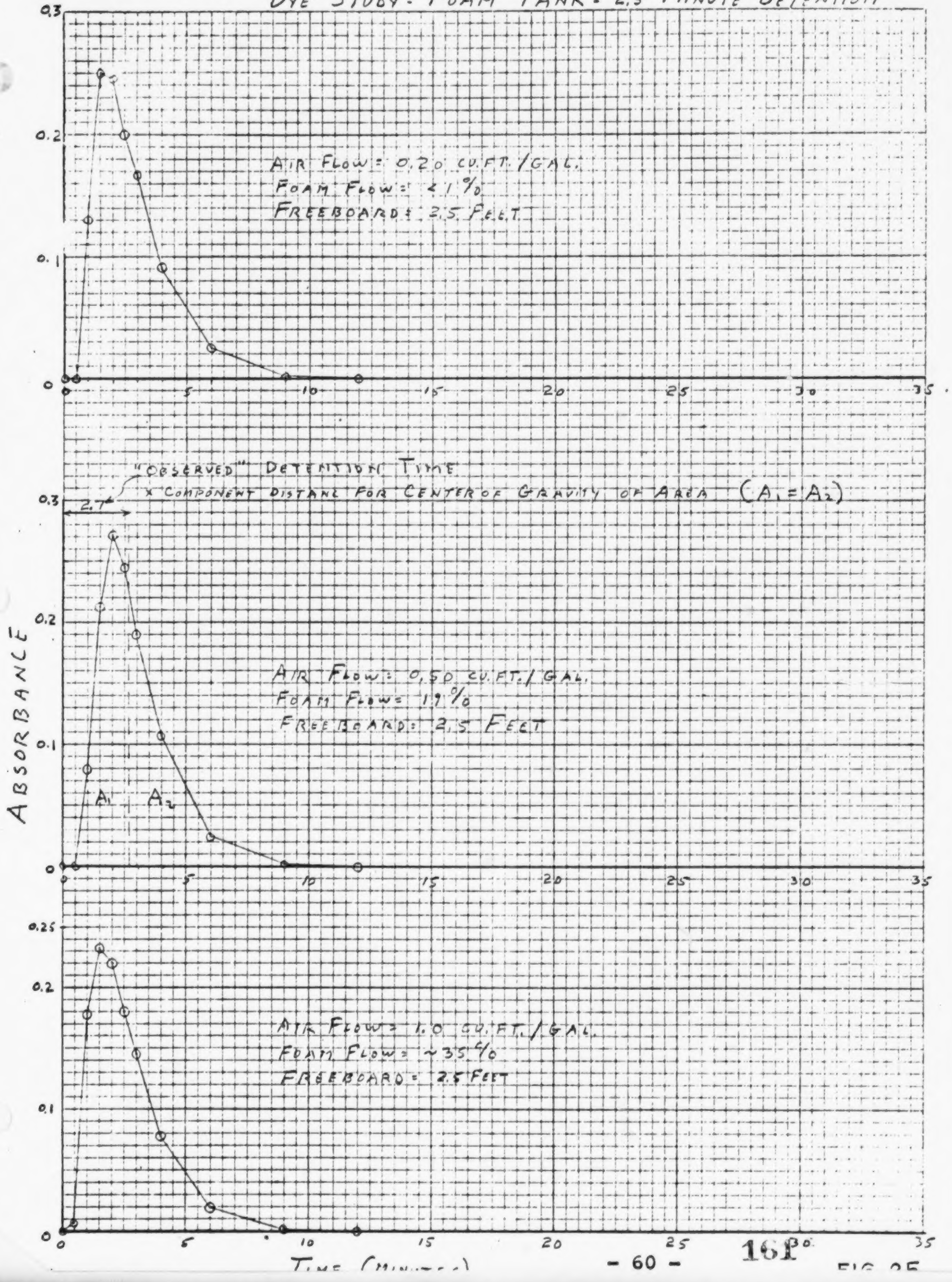


FIG. 24

Figure 25 2.5

DYE STUDY - FOAM TANK - 2.5 MINUTE DETENTION

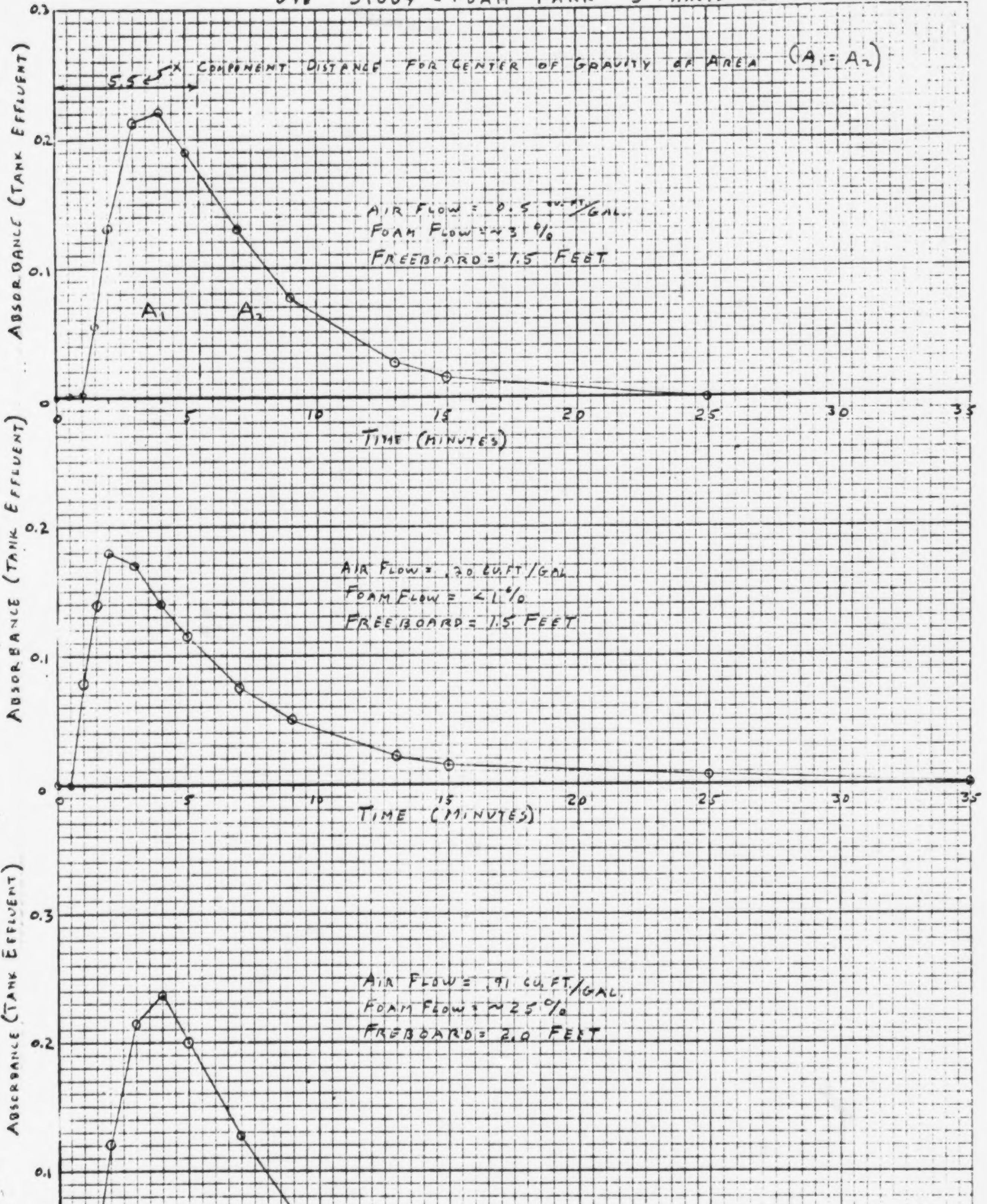


16P

FIG. 25

Figure 26

DYE STUDY - FOAM TANK - 5 MINUTE DETENTION



CLEARPRINT PAPER CO. NO. C3180. 10 DIVISIONS PER INCH BOTH WAYS 70 X 100 DIVISIONS. PRINTED IN U.S.A. ON CLEARPRINT TECHNICAL PAPER NO. 10

Figure 27

Dye Study - Foam Tank - 10 Minute Detention

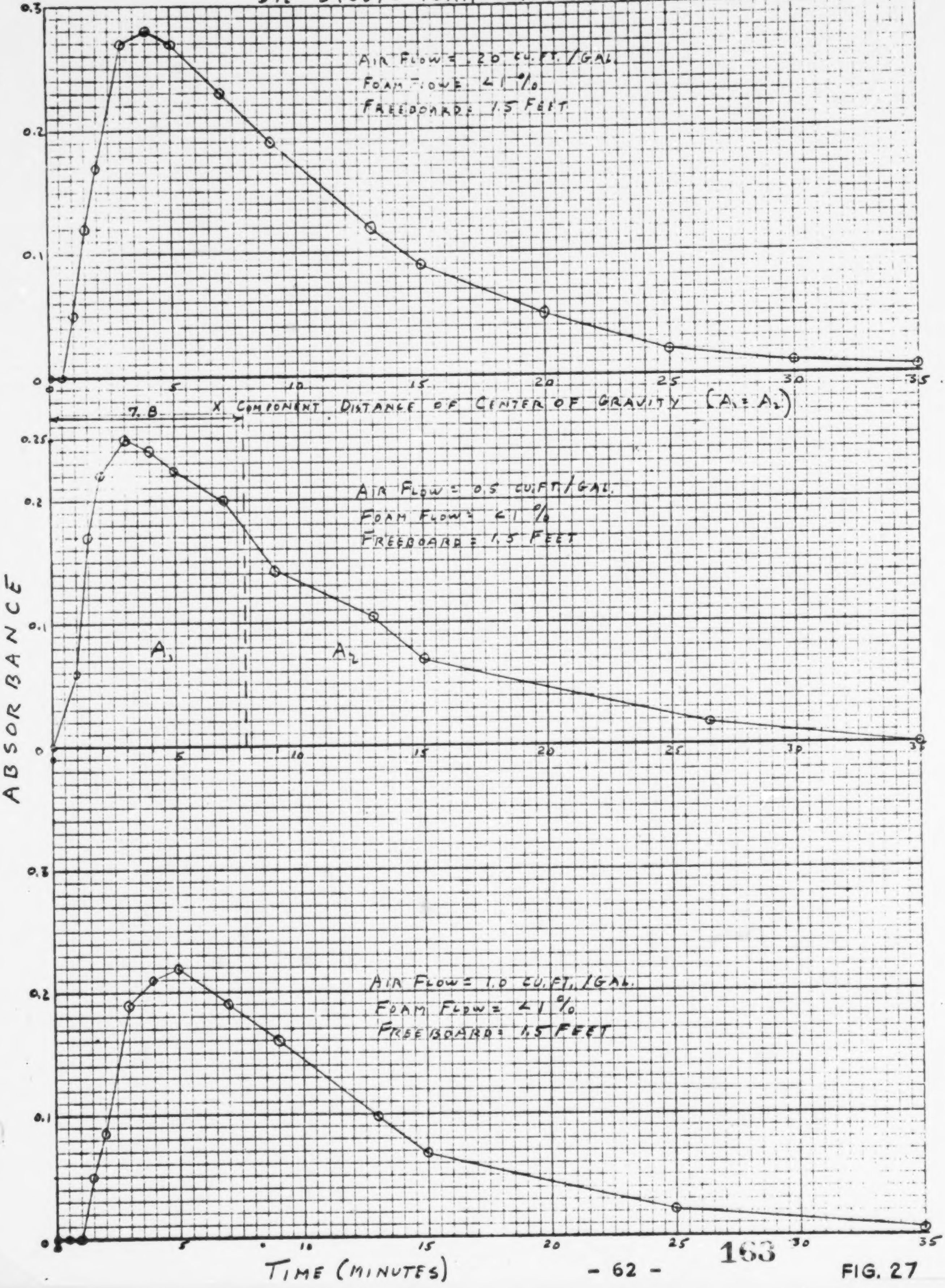
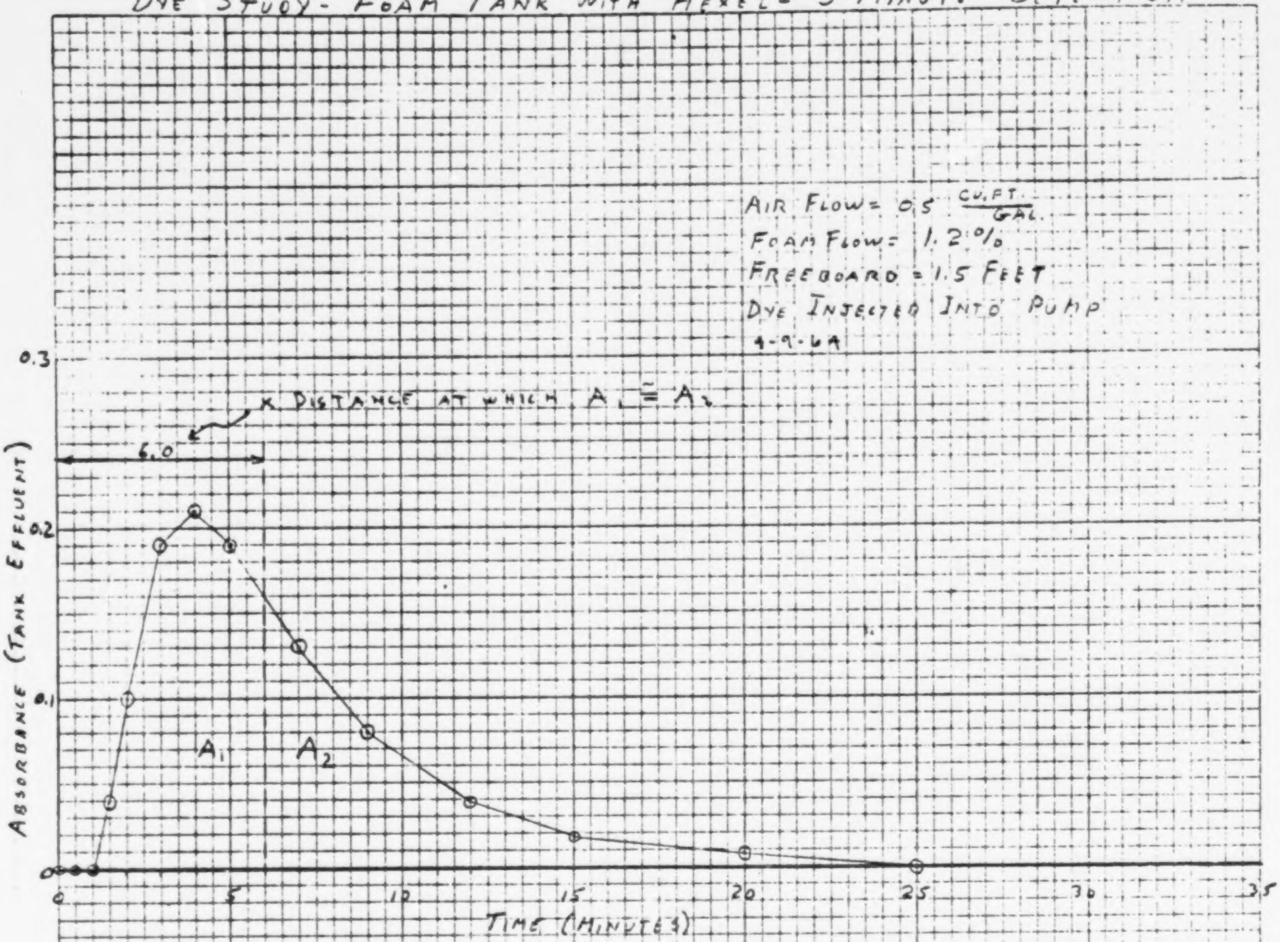


FIG. 27

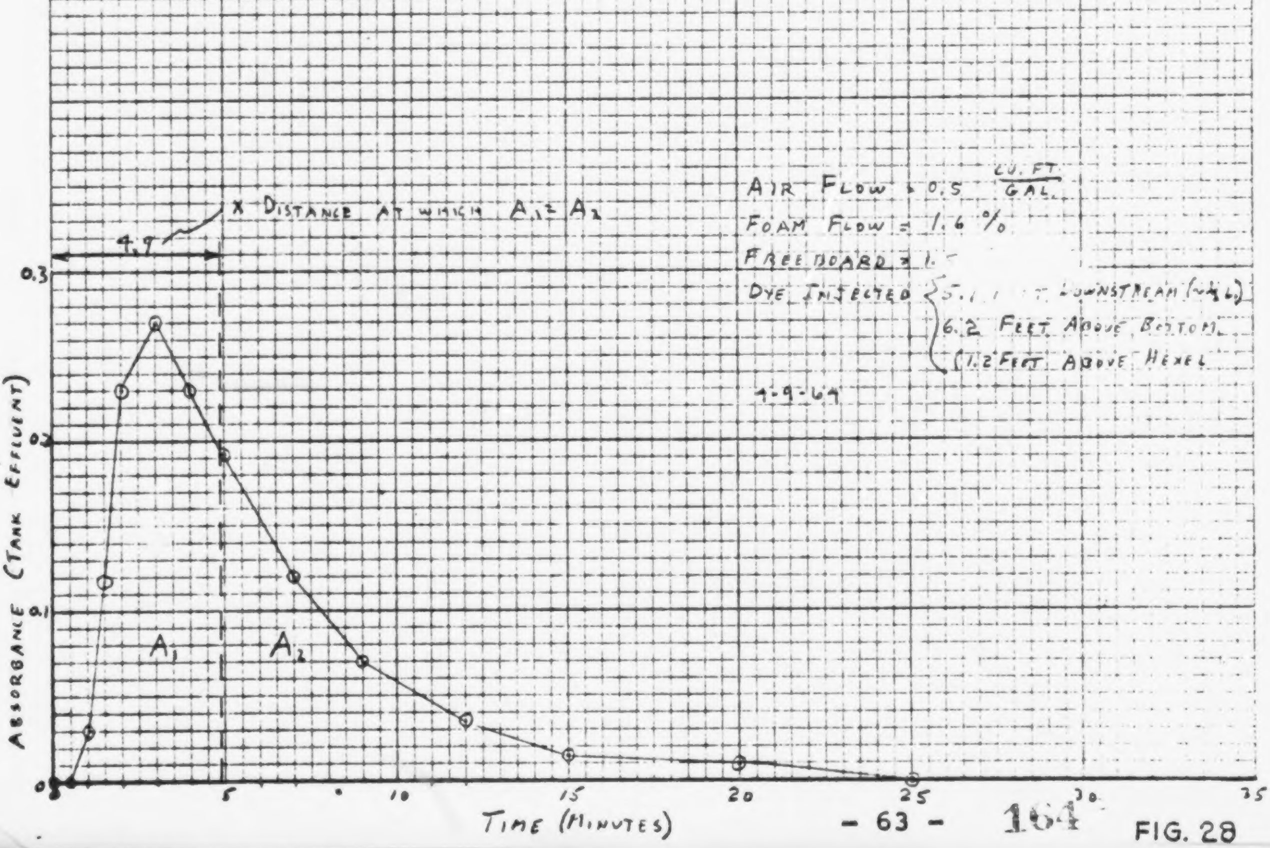
Figure 28

DYE STUDY - FOAM TANK WITH HEXEL - 5 MINUTE DETENTION



X DISTANCE AT WHICH A1 = A2

AIR FLOW = 0.5 CU.FT./GAL.
FOAM FLOW = 1.2 %
FREEBOARD = 1.5 FEET
DYE INJECTED INTO PUMP
4-9-6A



X DISTANCE AT WHICH A1 = A2

AIR FLOW = 0.5 CU. FT./GAL.
FOAM FLOW = 1.6 %
FREEBOARD = 1.
DYE INJECTED { 5.1 [illegible] DOWNSTREAM (~¼ L)
6.2 FEET ABOVE BOTTOM
1.2 FEET ABOVE HEXEL
4-9-64

FIG. 28

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Figure 29 - SS - COD

REMOVAL OF SUSPENDED SOLIDS & SUSPENDED SOLIDS COD

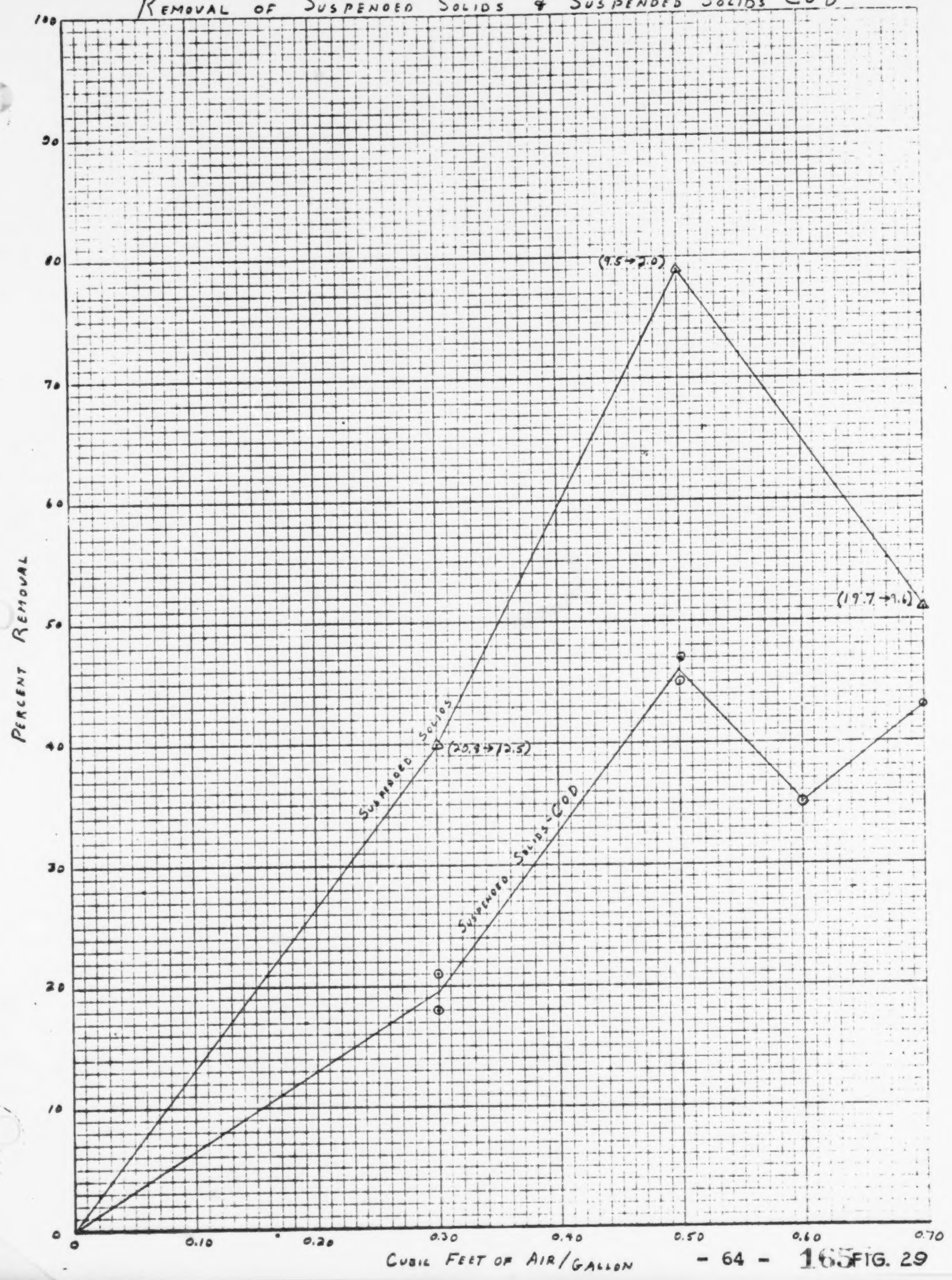
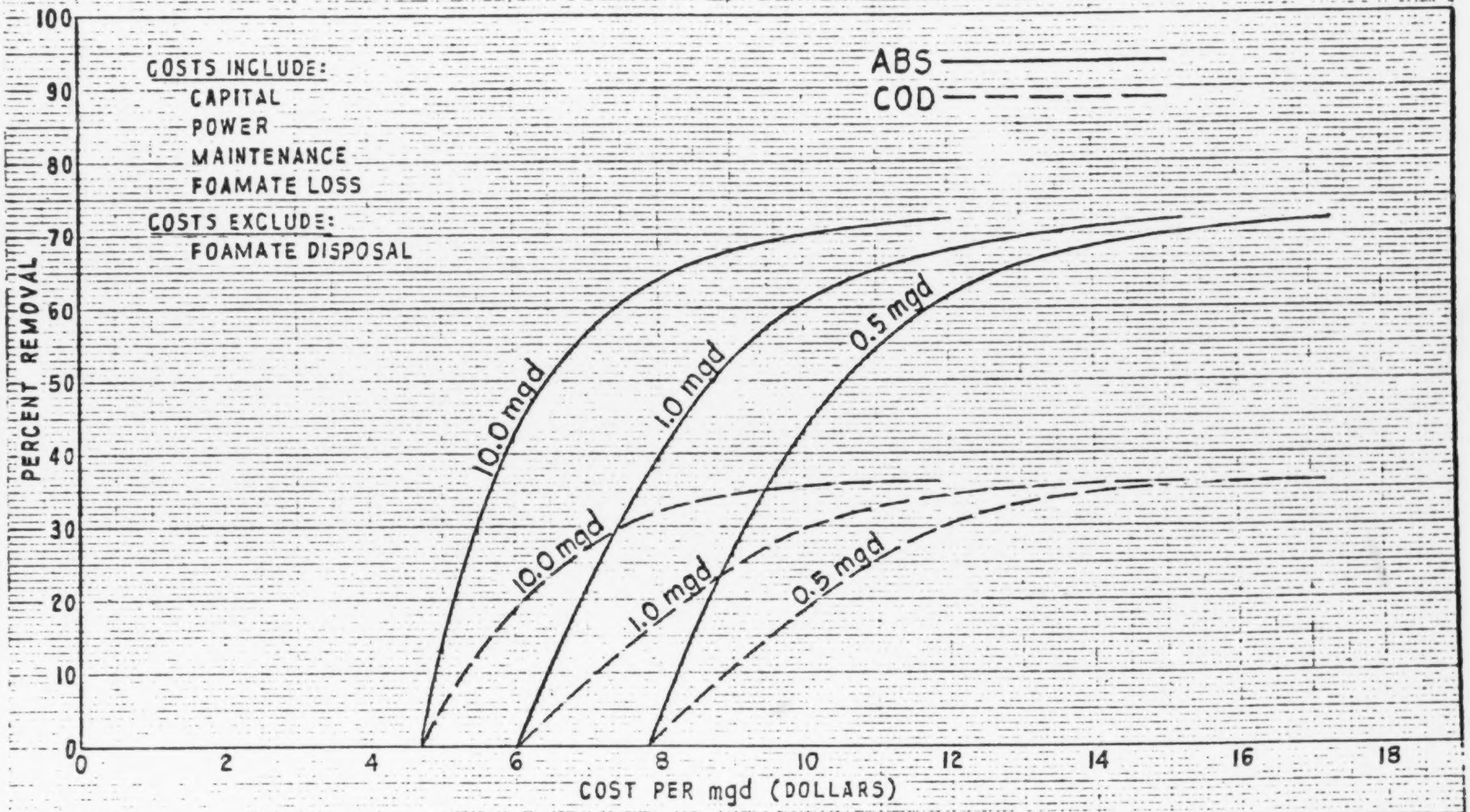


FIG. 29



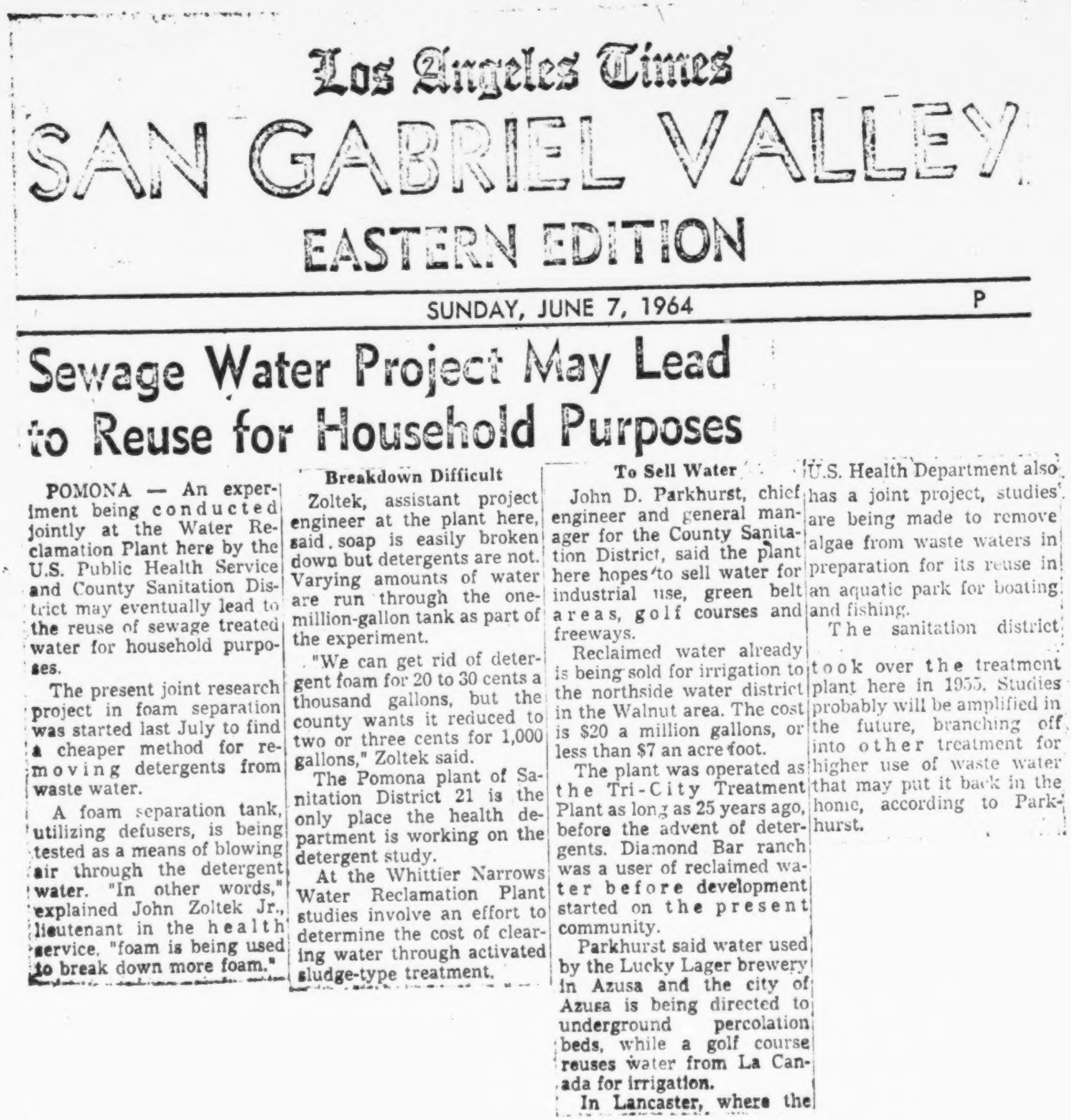
UNIT COST OF FOAM SEPARATION vs. REMOVAL EFFICIENCY

FIG. 30

TABLE 1

INVESTIGATION OF COD BALANCE

Source of Data (1)	COD Removed (MG/L) $(IN)_{COD} - (EFF)_{COD}$ (2)	Total COD Removed (5) + (6) (3)	$\frac{(3)}{(2)}$ as % (4)	SS - COD Removed (5)	(ABS) COD Removed (2 × Δ ABS) (6)	Necessary $(COD)_{ABS}$ (2) = (3) (7)	Necessary $\frac{mg\ COD}{mg\ ABS}$ For (2) = (3) (8)
Pomona 4-16-64	20.3	12.4	61	8.1	4.3	12.2	5.7
Pomona 5-7-64	8.3	5.8	70	2.4	3.4	5.9	3.5
Pomona 5-7-64	16.8	13.5	80	7.7	5.8	9.1	3.1
Pomona 5-12-64	13.7	11.4	83	6.4	5.0	7.3	2.9
Pomona 5-13-64	16.7	12.5	75	6.6	5.9	10.1	3.4
Pomona 5-18-64	25.9	19.7	76	9.7	10.0	16.2	3.2
Pomona 5-19-64	21.9	18.1	83	8.4	9.7	13.5	2.7



Los Angeles Times

SAN GABRIEL VALLEY

EASTERN EDITION

SUNDAY, JUNE 7, 1964 P

Sewage Water Project May Lead to Reuse for Household Purposes

POMONA — An experiment being conducted jointly at the Water Reclamation Plant here by the U.S. Public Health Service and County Sanitation District may eventually lead to the reuse of sewage treated water for household purposes.

The present joint research project in foam separation was started last July to find a cheaper method for removing detergents from waste water.

A foam separation tank, utilizing defusers, is being tested as a means of blowing air through the detergent water. "In other words," explained John Zoltek Jr., lieutenant in the health service, "foam is being used to break down more foam."

Breakdown Difficult

Zoltek, assistant project engineer at the plant here, said soap is easily broken down but detergents are not. Varying amounts of water are run through the one-million-gallon tank as part of the experiment.

"We can get rid of detergent foam for 20 to 30 cents a thousand gallons, but the county wants it reduced to two or three cents for 1,000 gallons," Zoltek said.

The Pomona plant of Sanitation District 21 is the only place the health department is working on the detergent study.

At the Whittier Narrows Water Reclamation Plant studies involve an effort to determine the cost of clearing water through activated sludge-type treatment.

To Sell Water

John D. Parkhurst, chief engineer and general manager for the County Sanitation District, said the plant here hopes to sell water for industrial use, green belt areas, golf courses and freeways.

Reclaimed water already is being sold for irrigation to the northside water district in the Walnut area. The cost is $20 a million gallons, or less than $7 an acre foot.

The plant was operated as the Tri-City Treatment Plant as long as 25 years ago, before the advent of detergents. Diamond Bar ranch was a user of reclaimed water before development started on the present community.

Parkhurst said water used by the Lucky Lager brewery in Azusa and the city of Azusa is being directed to underground percolation beds, while a golf course reuses water from La Canada for irrigation.

In Lancaster, where the

U.S. Health Department also has a joint project, studies are being made to remove algae from waste waters in preparation for its reuse in an aquatic park for boating and fishing.

The sanitation district

took over the treatment plant here in 1955. Studies probably will be amplified in the future, branching off into other treatment for higher use of waste water that may put it back in the home, according to Parkhurst.

Article on
The Joint CSD-PHS Foam Separation Study

Los Angeles Times - 7 June 1964

Joint PHS-LACSD Water Renovation Research Project

PART THREE

COST-QUALITY STUDY

Whittier Narrows

Water Reclamation Plant

Franklin D. Dryden - Project Engineer

Cost-Quality Study

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PROGRESS REPORT ON
COST VERSUS QUALITY STUDIES

Chapter 1

INTRODUCTION

a. Object

The object of this study is to develop a relationship between the quality of a renovated water and the cost of producing it by the standard activated sludge, contact stabilization, and step aeration processes.

b. Experimental Approach

The plan for meeting the above objective was to operate a large activated sludge plant utilizing in sequence, each of the three processes. Operations would be initiated for each of the processes at a flow rate of approximately the nominal plant capacity of 10 mgd. The feed rate to the plant would then be incrementally increased until the effluent quality deteriorated to a level considered unacceptable by the parties involved. Since the maximum capacity of the two influent pumps is 18.5 mgd, it will probably be impossible to push average daily flow to the experimental plant above 17 mgd.

c. Location and Description of Research Facility

The study was conducted at the Sanitation Districts' Whittier Narrows Water Reclamation Plant (Figure 1), located at 301 N. Rosemead Blvd., El Monte, and within the reservoir area behind Whittier Narrows Dam. The Whittier Narrows Dam is only used for flow control purposes; the ground at the plant site will only be inundated by a storm of once-in-30-years intensity.

The Whittier Narrows Water Reclamation Plant was placed in operation on August 20, 1962. The plant was constructed primarily to demonstrate on a large scale, the feasibility of waste water renovation and was built solely for the water it would produce. The water is delivered to spreading basins located south of the Whittier Narrows Dam.

The plant receives suction from a 60" trunk sewer with an average 50 mgd flow predominantly composed of domestic sewage from the San Gabriel Valley. One of the two influent pumps is fixed speed and the other is equipped with a variable speed drive. Minimum flow in the trunk sewer is approximately 18 mgd. The capacity of either pump alone (11 mgd) exceeds the basic design capacity of the plant by the standard activated sludge process. Under normal operation, the plant flow can be maintained at a constant 24-hour rate.

In order to have an opportunity to produce the best possible water at the lowest possible price, the plant was designed with considerable flexibility. Primary effluent may be metered and introduced at the quarter points along each aeration tank, and return activated sludge can be metered and introduced at the head end of each tank and at the foot of the middle tank. The pumping of return activated sludge is automatically controlled to maintain a preset sludge blanket level by the use of a turbidimeter. Three centrifugal air blowers provide an air capacity range of 5,000 to 27,000 cfm. The air distribution within each tank may be varied by changing the number and arrangement of spargers along the swing air headers. This flexibility permits operation of the plant by the three flow configurations outlined in the project objective.

There were several problems which interfered with steady state operation of the plant during the study period. The influent pumps were not supplied in accordance with the design specifications and exhibited the undesirable trait of ragging up 2 or 3 times a day. Since there is no operator in the plant during most of the night, the ragging resulted in irregular curtailment of flows. The redesigned pump impellors were installed in late 1963 and minor ragging now occurs on the average of one pump per day.

The air supplied to the aeration tanks is withdrawn from under the covers on the primary sedimentation tanks and channels to exercise positive odor control. At high flow rates, the water level in distribution channel No. 2 under the blower line suction port was raised sufficiently to cause a virtual water spout with excessive evaporation of water which then condensed in the air lines and blowers. So much was evaporated at high air rates that only 2 of the blowers could be operated at one time without filling the suction line with water. This condition was relieved in April when a floating horizontal plate was installed under the suction port thereby preventing evaporation at the point of highest air velocities. This change allowed the increase in air rates that occurred on May 5th and also was instrumental in permitting increases in plant flow to the September rates of 16 mgd.

The plant effluent is metered through a water works type propeller meter which had given generally satisfactory results according to previous checks but which appeared to be erroneous in June. Whatever might have caused the irregular results obtained at that time, it was apparently solved by the combined effects of removing and checking the meter and installing a liner in the metered discharge line in order to increase the average velocity. Judging by the data contained in this report, it is unlikely that the meter was significantly in error during the major portion of the study period. It now checks very well and should give completely reliable results for the rest of the study.

Occasional power failures and mechanical failures such as any large plant experiences also contributed to deviations in plant flow and loading ratios with resulting reactions in the SVI and to a minor extent in effluent quality parameters..

d. Sampling Methods and Equipment

In addition to the regular sampling and analysis performed at the Whittier Narrows plant to assist in process control, 24-hour composited refrigerated samples of the plant influent, primary effluent, return activated sludge, secondary effluent, and final effluent were collected weekly. The analysis schedule is presented in Figure 2. Where there are dark boxes on the data sheet, no analysis or observation was made, the clear boxes were filled in by the plant laboratory technician, and the data in the shaded boxes was obtained from other plant records or from analysis of the samples at the Districts' Central Laboratory.

Samples were automatically composited with a timer controlled rotating scoop dip sampler supplied by the Robert A. Taft Sanitary Engineering Center. The scoop operated at 15-minute intervals, dipping a sample from a 2" wide, 4" deep and 24" long continuous dip trough and depositing the sample into a 1-gallon container inside a small refrigerator. The continuous flow sample was obtained with a small Moyno progressing cavity, positive displacement pump (Figure 3). The automatic samplers were installed and operating by the middle of January; the first weekly 24-hour composite was collected between 8:00 am, January 20 and 8:00 am, January 21, 1964. On a few occassions, no samples were collected due to mechanical failure or unrepresentative conditions at the plant which resulted in meaningless samples.

Chapter 2

PROCESS DESCRIPTIONS

a. Definitions

The modifications of the activated sludge process of interest to a discussion of results of the work reported here are: standard activated sludge, contact stabilization, and step aeration. Each of these processes is defined as follows:

Standard activated sludge is a process in which all the return sludge is mixed with all the primary effluent at the head end of the aeration tanks. In the case of Whittier Narrows, the three 300'-long aeration tanks may be operated in this manner either in series or in parallel without distinction.

Contact stabilization provides aeration of all returned sludge in a separate tank for several hours preceding its introduction to the mixed liquor aeration tank. Mixed liquor aeration time is usually 1-1/2 to 3 hours.

Step aeration is a system of introducing the primary effluent at several locations along the aeration tank usually including the head end. At Whittier Narrows step aeration can be achieved by distributing primary effluent among the three aeration tanks, among two of the tanks, or at quarter points along the length of one to three tanks. Special studies of mixing in each tank indicated that longitudinal mixing is so pronounced that the performance of any one aeration basin is not significantly affected by the point of introduction of the flow although the average detention time is affected. Step aeration within a single aeration tank at Whittier Narrows is not considered a process variable for the purposes of process evaluation and only when the inflow is distributed among Aerator No. 1 and No. 2 is the term applied.

b. Process Schedule

The types of processes employed during the period of this report are indicated in the bottom graph of Figure 19 which shows the Distribution of Primary Effluent to Aerator No. 1. Aeration tank No. 1 also received all return activated sludge during this reporting period. The primary effluent not introduced to Aerator No. 1 was stepped into Aerator No. 2. Separate aeration of the return sludge alone is the distinguishing feature of the contact stabilization process; whereas the distribution of primary effluent among the basins is a distinguishing feature of the step aeration process. Examination of Figure 19 shows that the portion of primary effluent going to Aerator No. 1 during two periods, mainly January, February, and most of March and from mid-April to mid-May was so negligible that the contact stabilization process may be considered to have been employed during these periods. In the remaining periods (mid-March to mid-April and mid-May to July 1), the percentage of primary effluent going to Aerator No. 1 ranged from 25 to 65% thereby placing the process in the step aeration catagory.

The reason for these changes is related to one of the methods employed at the plant for controlling sludge settleability, and thereby the sludge volume index. Increasing reaeration time for the return sludge makes it more dense and settleable; conversely, shortening reaeration time makes it less dense. Changes in reaeration time can be accomplished by increasing or decreasing the amount of primary effluent introduced into the first aerator.

Reaeration time is also influenced by changes in sludge settleability. The rate of sludge return is automated by the use of turbidity meters to maintain a constant level for the sludge-supernatent interface in the final sedimentation tanks; therefore, the more dense the sludge, the lower the return rate. The automatic return sludge control tends to reinforce any change in sludge characteristics engendered by changing plant flow, plant air rates, or distribution of primary effluent to the first aerator; a reduction in return sludge flow rate resulting from an increase in sludge density works to further increase sludge density and the process moves in the inverse direction as sludge densities decrease.

It should be noted that contact stabilization can be considered a special case in the spectrum of possible distribution arrangements conceivable under a concept of step aeration which provides that primary effluent can be distributed at any point along the aeration tank. The choice of distribution patterns is related to plant flow, temperatures, sludge characteristics, air distribution patterns, and desired results. This explains the irregular alternation between, and the blending of, the two processes during this period of operation.

During the summer and fall of 1964, successful operation at maximum flow rates using the step aeration process was accomplished. In the near future the process will be converted to standard activated sludge and after about a year's operation to contact stabilization; in both cases process control will be accomplished by varying air and plant flow rates but without changing points of distribution of primary effluent.

Chapter 3

DISCUSSION OF DATA

a. General Trends

In broad terms, there was a continuous increase in plant flow from about 10 mgd in February to about 13 mgd in June (Figure 4). The average sewage temperature also increased steadily from 68° F in January to 76° F in late June (Figure 4). During this same period, there was a gradual decrease in the suspended solids (SS), COD, and BOD of the raw sewage and of the primary effluent (Figures 5, 7 and 9). Organic nitrogen was more erratic but also showed a downward trend (Figure 11). Although the steady decline could be related to an annual increase in per capita water use or decrease in per capita discharge of waste products, it is more probably the result of the steady increase in sewage temperatures which supported a greater biological activity and a consequent reduction of BOD in the sewers. This trend, and the higher levels of biological activity within the activated sludge tanks with increasing temperatures, accounts for the plant's ability to treat 30% more sewage in June than in February. Throughout the 6 months, there was a relatively uniform 90% reduction in COD and BOD (Figures 6 and 8) and about a 95% reduction in suspended solids (Figure 10) through the plant. Organic nitrogen removal stayed at about 80% (Figure 12). Dissolved oxygen levels in the effluent ranged between 0.5 and 1.5 mg/l (Figure 14) ahead of the effluent weir in the final tank.

Superimposed on the general trends described above were the fluctuations in the sludge volume index (SVI), the effluent quality as measured by SS, BOD, COD, nitrate and turbidity, the air rate, the mixed liquor volatile suspended solids (MLVSS), the aerator loading, and the distribution of primary effluent into the first aerator. Undesirable changes in the SVI and the effluent quality parameters were countered by changing the air rate, the plant flow, or the distribution of primary effluent as discussed in later sections.

b. Effluent Quality

As indicators of effluent quality, the secondary effluent suspended solids, chemical oxygen demand (COD), and turbidity have proved most satisfactory and consistent, (Figures 13 and 15). Biochemical oxygen demand (BOD) does not always follow the other three (Figure 13) because of the periodic influence of nitrifying cultures which produce nitrates in the BOD bottles and thereby exert a high oxygen demand without relation to the true quality of the effluent. Nitrates, however, can influence effluent quality by releasing nitrogen gas in the sludge blanket in the final sedimentation tanks and buoying sludge to the surface. Such rising sludge has never been experienced at Whittier Narrows.

Throughout most of the six month period for which data was obtained, there is good agreement between at least two of the three basic indicators of effluent quality. Peaks in suspended solids occurred on February 10, April 7, and May 5 and peaks of effluent COD occurred on the same dates (Figure 13). Turbidity measurements by both the Secchi Disk and Turbidimeter readings (Figure 15) check the SS and COD peaks of February and April but are not so consistent in May. COD peaks on March 17 and June 30 do not coincide with either suspended solids or turbidity readings, but a COD peak on May 25 is supported by deterioration in turbidities on the same date and a marked improvement in COD on June 9 matches an improvement in turbidity.

It should be noted that the turbidity measurements plotted in Figure 15 are mirror image because a deterioration in effluent quality results in an increasing turbidity by the Jackson Candle Standard Turbidity Unit employed in the Hach turbidimeter but a decrease in the distance a Secchi Disc may be seen when suspended in the effluent. With one or two exceptions, the two curves correlate very closely.

c. Nitrate BOD

If the BOD test is performed without sterilizing by heat or acidification, the organisms not only exert a carbonaceous oxygen demand but if nitrifying bacteria are present oxygen will be used to form nitrates. Nitrification tends to occur in the plant when there is an abundant supply of air. As can be seen in Figure 17, the air supplied per gallon of influent was increased in early February by decreasing the plant flow rate (Figure 3). At this time the growth of nitrifying bacteria was apparently stimulated and nitrate in the effluent took a sudden rise on February 17 followed by an erratic but generally steady decline until it had returned on April 1st to an even lower value than it started. The decline in nitrate is generally matched by a steady decline in the quantity of air per gallon over the same period. The most interesting relationship, however, is the almost exact correlation of every peak and valley in the nitrate curve during the two month period with the peaks and valleys of the BOD curve offset by one week. It would appear that during the 5 days when the nitrate was developing in the BOD bottle and thereby exerting an excess BOD it was also developing in the same way in the microflora of the activated sludge. It should be noted that BOD's are affected by other factors besides nitrates and the correlation is not so clear in the remainder of the year. However, as nitrification began to occur again about May 10th (as the result of increasing the air supply) there was also an increase in BOD in spite of very low suspended solids values and a decided drop in the COD.

d. The Sludge Volume Index (SVI)

The sludge volume index is an attempt to describe the settleability of the activated sludge solids in the mixed liquor in relation to the concentration of these solids. It is obtained by dividing the volume of solids obtained during 30-minutes of settling in a 1 liter cylinder expressed in percent by the suspended solids concentration expressed in percent. Unfortunately, if nitrates are present in significant concentrations the oxygen in the nitrate will be consumed by the bacteria in the sludge thereby releasing nitrogen gas. If the test is performed in a 1 liter cylinder in the laboratory, the gas occupies an appreciably greater volume than it does if the sludge is at the bottom of the final aeration tank at approximately 4 psi pressure. The volume the gas occupies not only contributes to higher settleability measurements by taking up space, but, since buoyancy is directly related to the volume of water displaced, prevents settling by exerting a greater uplift. Because the object of the SVI test is to simulate what is happening to the sludge in the final tank, in June the Sanitation Districts instituted the practice of performing the settleability test by subjecting the test cylinder to a 4 psi pressure whenever nitrification appeared to be affecting results. This procedure provides excellent correlation between the percent the return sludge flow comprises of the flow entering the final sedimentation tanks.

The sludge volume index can be decreased or increased by making changes in the plant flow rate, the air supplied per gallon, or the length of reaeration time. In each case, a change which increases the amount of air reaching the sludge during one cycle through the tanks will tend to decrease the SVI by increasing the ash content of the sludge through destruction of volatile matter thereby making the sludge heavier in relation to solids concentration; lowering the air-solids ratio will increase the SVI. If the SVI is too high, the sludge does not settle easily and some unsettled floc will be carried over the weirs. If the SVI is too low the effluent quality appears to deteriorate. This is thought to result from the major portion of the sludge settling too rapidly and thereby, not enmeshing particulate matter in its flocculant masses.

Two of the three periods previously described as being most consistently poor in effluent quality, February 10 and May 5, occurred when the SVI was in excess of 140 but on April 7 and the SVI was only 104. However, on April 7 the plant flow was raised to 14 mgd for the first time and the deterioration in effluent quality was probably related to the high flow even though the SVI had not yet been affected.

During the period between January and March, the suspended solids of the mixed liquor (Aerator No. 3) were relatively constant (Figure 20) and the fluctuations in the SVI were related to changes in settleability and therefore in part to the nitrate pattern as described previously. From March through June the trends in SVI can be well correlated with changes in the distribution of primary effluent (Figure 19) and an increase in the air rate on May 5 (Figure 17). The general lowering of the SVI for this period is probably related to the somewhat higher level of suspended solids carried in the mixed liquor, but the distinct periods of rise and fall can clearly be related to the transfer of primary effluent into or out of Aerator No. 1 (Figure 19). Increasing the amount of primary effluent introduced into Aerator No. 1 results in a gradual increase in the SVI and decreasing the percentage produces the opposite effect. The increase in air rate that occurred May 5 coupled with virtually no primary effluent into Aerator No. 1 produced a rapid decline in the SVI. Gradual step increases in the percent of primary effluent being diverted into the No. 1 tank which commenced May 15th were not able to reverse the downward trend until about May 30. It would appear that changes in air rates produce more rapid changes in the SVI because all the sludge is immediately subjected to an increase in the air rate, whereas changes in flow rates and distribution patterns have a gradual effect as the sludge recirculates and changes in character. The higher temperatures since April may also account for a more rapid response in the process to changes in variables, whereas in the winter it is not apparent that the relatively greater increase in air rate accomplished by changing the plant flow in February resulted in as rapid or as related a change in SVI as occurred in May.

The general mechanism by which either holding the plant flow constant and changing the distribution of primary effluent into the aeration tanks or holding the distribution constant and varying the plant flow affects the sludge volume index is quite similar. Increasing the percent of primary effluent distributed to Tank No. 1 decreased the reaeration time and therefore decreases the amount of air per pound of volatile suspended solids per pass through the tank. As less organic material is destroyed, the concentration of suspended solids in the mixed liquor slowly increases and the higher volatile content impairs its settling properties. A greater volume of sludge is therefore created and as it is returned to the aeration tanks the detention time in the tanks is further reduced. Increasing the plant flow has the same effect by putting more load on the tank, creating an increase in suspended matter and an increase in the sludge return rate with a corresponding decrease in reaeration time.

e. Future Plans

Every effort commensurate with the requirement to maintain plant effluent quality will be made by the plant operators to avoid changing more than one of the control parameters at a time. During the study period of this report there was no time during which simultaneous changes were made which obscured their related effects. Changes in flow can, of course, occur due to mechanical failure but it is unlikely that such accidents did or will materially influence results.

In order to provide a more frequent sampling of influent quality, daily samples will be air mailed to the Taft Center for Total Carbon analysis. It is felt that the week between sampling permits possibly significant changes in sewage character to avoid detection.

In the near future the process will be converted to standard activated sludge and after about another year's operation to contact stabilization. Late summer operation at 16 mgd by the step aeration process has been accomplished without deterioration in effluent quality. Since the maximum capacity of the two influent pumps is 18.5 mgd it will probably be impossible to push average daily flow above 17 mgd.

Chapter 4

COSTS

a. Capital Costs

Capital costs for the Whittier Narrows Water Reclamation plant are based on the actual construction contract price plus 7% for engineering. In Table 1, the contract price of approximately $1,352,000 has been proportioned among the basic plant components. The primary sedimentation tank, aeration tanks, final sedimentation tanks, and blowers are considered to be directly related to the quality of treatment. The cost of influent pumps, the control building, the chlorine contact chambers, and site preparation are not related to secondary effluent quality and therefore will not be included in the capital cost evaluation.

The value of the units used in treatment, $1,045,000, has been amortized over 25 years at 4% interest to obtain a uniform annual cost. The uniform annual cost of $67,000 divided by the annual water production in millions of gallons for any flow rate gives the capital cost per million gallons at that flow rate. The curve for capital cost per million gallons of secondary effluent presented in Figure 23 is based on using all primary, aeration and final tanks. In the coming year's study, the plant will be operated with one or more of each type of tank out of service to provide a basis for relating the effluent quality to the cost of those units in service.

For the months of January and February, two primary sedimentation tanks were in service, but between March and June only one was used. At the flow rates encountered during the period there are few differences in average effluent quality (Table 2). At flows under 13 mgd and at any but the lowest annual temperatures, the use of only one primary does not seem to have any adverse effects.

b. Labor Costs

Labor costs are related to wage scales in different localities and are therefore difficult to generalize upon. For this reason the labor is first presented in terms of hours per million gallons and then the actual labor rates of the men employed at Whittier Narrows are used to determine the cost curve applicable during this study. Both methods of relating labor costs are used in Figure 24.

Three men--two operators and a laboratory technician--comprise the operating force at Whittier Narrows. Doubling the size of the plant would probably require an additional operator. As plants increase in size the relationship between labor and flow results in a saw-tooth curve with an upward step in the unit cost as an additional man is required followed by a gradual decline until the increase in plant capacity justifies another man.

Unusual maintenance labor requirements are not included in the costs since they have no relation to flow or to how the plant is being operated. At Whittier Narrows, such maintenance work is performed by a Sanitation District Maintenance crew, whereas regular maintenance is performed by the operators. The three men employed at Whittier Narrows each work 44 hours per week and have an average wage of $3.60 per man-hour. The average daily labor cost is therefore $68.00 per day regardless of flow or effluent quality. At 10 mgd, the labor cost would equal $6.80 per million gallons.

c. Power

In addition to direct labor, power for operation of blowers is the only other significant operating cost. The power required for influent pumping is peculiar to the physical location of the Whittier Narrows plant and has no relation to effluent quality.

The three centrifugal blowers at Whittier Narrows use horsepower in proportion to the quantity of air produced. Unlike positive displacement blowers, the centrifugal blower can be throttled with only a slight change in efficiency and there is no power wasted in the form of surplus air released to the atmosphere. Two of the three blowers produce 11,000 cfm with approximately 400 hp and the third puts out approximately 5,500 cfm with 200 hp. Assuming 90% efficiency and a power cost of $0.0085 per KWH, the cost of air is $4.25 per million cubic feet of air (MCF). Since the cost per million cubic feet is relatively constant and since the portion of the total air production going to the aeration tanks is metered and recorded, the cost of air follows the same curve as the unit rate of air in cu.ft./gal plotted in Figure 17. As shown in the total air supply curve of Figure 17, the actual air rate was relatively constant between January and May at which time the third blower was placed in operation. Since there was a relatively gradual and continuous increase in plant flow during the period, the unit air rate decreased from 2.35 cu.ft./gal in February to a low of 1.5 cu.ft./gal in April. In May, it again exceeded 2.0 cu.ft./gal as the air was increased and dropped to about 1.8 cu.ft./gal toward the end of June. Since May the unit air rate has hovered around or slightly below the 2.0 cu.ft./gal rate. At 2.0 cu.ft./gal the cost of air is $8.50 per million gallons.

d. Future Studies

At this time, there is not sufficient data to draw relationships between effluent quality and costs. The key parameters are the flow and the air supply per gallon of flow. During the first six months the range of flow went between 9 and 13 mgd but in August 1964 it has already averaged 16 mgd. There have been times when both good and poor quality water have been produced at all flow rates. Usually the period of time in which combinations of low flow and high air rate existed were relatively short (February) and it is likely that the change contributed to the subsequent improvement in effluent quality although the nature of the time lag is not certain. Later in the study it is anticipated that there will be enough information for periods of extended consistent operation to support valid cost versus quality relationships under stable operating conditions.

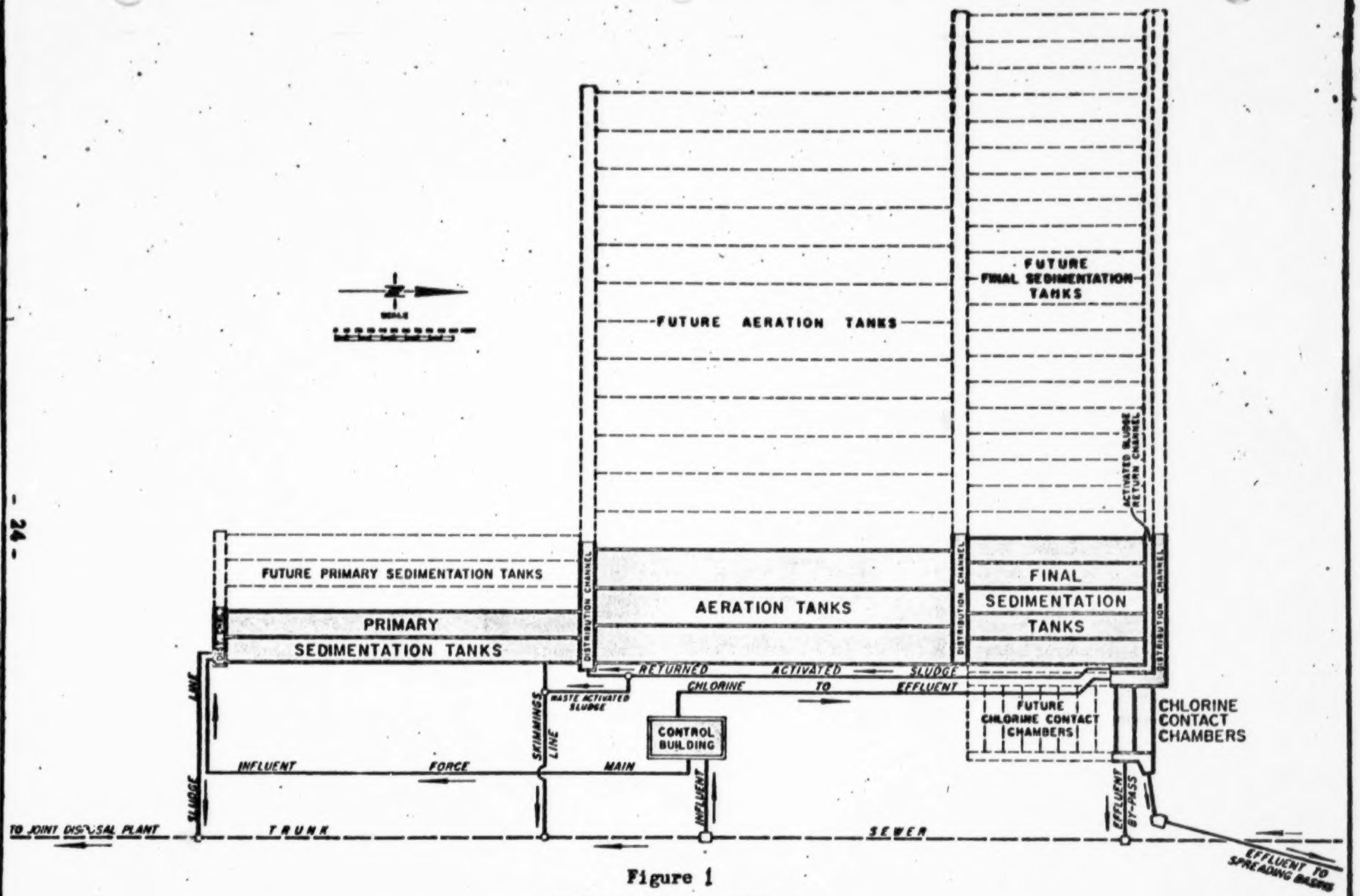

FUTURE FINAL SEDIMENTATION TANKS
FUTURE AERATION TANKS
ACTIVATED SLUDGE RETURN CHANNEL
FUTURE PRIMARY SEDIMENTATION TANKS
PRIMARY SEDIMENTATION TANKS
DISTRIBUTION CHANNEL
AERATION TANKS
DISTRIBUTION CHANNEL
FINAL SEDIMENTATION TANKS
DISTRIBUTION CHANNEL
RETURNED ACTIVATED SLUDGE
CHLORINE TO EFFLUENT
WASTE ACTIVATED SLUDGE
FUTURE CHLORINE CONTACT CHAMBERS
CHLORINE CONTACT CHAMBERS
CONTROL BUILDING
SLUDGE LINE
INFLUENT FORCE MAIN
SKIMMINGS LINE
INFLUENT
EFFLUENT BY-PASS
TO JOINT DISPOSAL PLANT
TRUNK SEWER
EFFLUENT TO SPREADING BASINS
N
SCALE

Figure 1

COUNTY SANITATION DISTRICTS
OF LOS ANGELES COUNTY, CALIF.
OFFICE OF CHIEF ENGINEER

WHITTIER NARROWS WATER RECLAMATION PLANT

JOHN D. PARKHURST - CHIEF ENGINEER

FIGURE 2

WEEKLY DATA SHEET

JOINT U.S.P.H.S - C.S.D. COST VS. QUALITY STUDY

WHITTIER NARROWS WATER RECLAMATION PLANT

DATE FROM ______
TO ______

OPERATOR ______

COMPOSITE HRS. ____

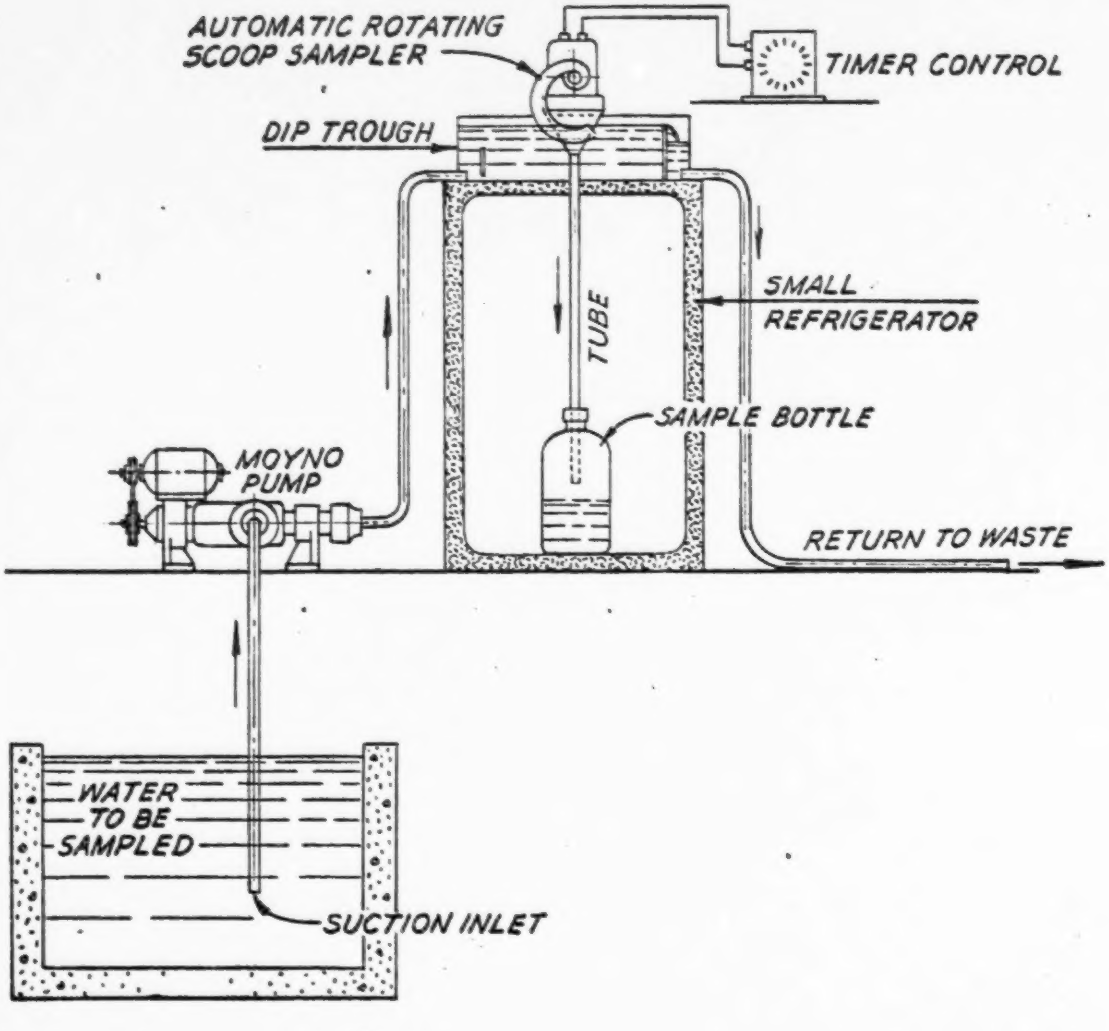
	FLOW MGD		SUSP. SOLIDS	C.O.D.	B.O.D. 5 DAY 20°C	A.B.S.	NH_4 AS N	NO_3 ① AS N	NO_2 ① AS N	N ORG. AS N	N TOTAL	PO_4	D.O. ①	CHLORINE		pH	TEMP.	TURBIDITY		COLOR	ODOR	
	WKLY AVG.	DAY SAMPLED	MG/L	MG/L	MG/L	MG/L	MG/L	MG/L	MG/L	MG/L	MG/L	MG/L	MG/L	REQUIRED MG/L	RESIDUAL MG/L		°F	SECCHI FT.	METER UNITS			
PLANT INFLUENT																						
PRIMARY EFFLUENT																						
RETURN ACTIVATED SLUDGE																						
SECONDARY EFFLUENT																						
FINAL EFFLUENT																						

		DAY SAMPLED	WEEKLY AVERAGE
WASTE ACTIVATED SLUDGE	Q-MGD		
A.B.S. - WASTE LINE	Q-MGD		
CHLORINE DOSE	LBS/DAY		
S.V.I.			
AIR	CFM/CF/GAL	/	/
PROCESS PATTERN			

		SUSPENDED SOLIDS		SETTLEABLE SOLIDS	DETENTION TIME
		MG/L	% VOL	ML/L - 30 MIN.	HOURS
AERATOR No	1				
	2				
	3				
RETURN SLUDGE					

REMARKS: ①SPOT CHECKED

2-370-a-10


AUTOMATIC ROTATING SCOOP SAMPLER
TIMER CONTROL
DIP TROUGH
SMALL REFRIGERATOR
TUBE
SAMPLE BOTTLE
MOYNO PUMP
RETURN TO WASTE
WATER TO BE SAMPLED
SUCTION INLET

TYPICAL AUTOMATIC SAMPLING ARRANGEMENT

FIG. 3

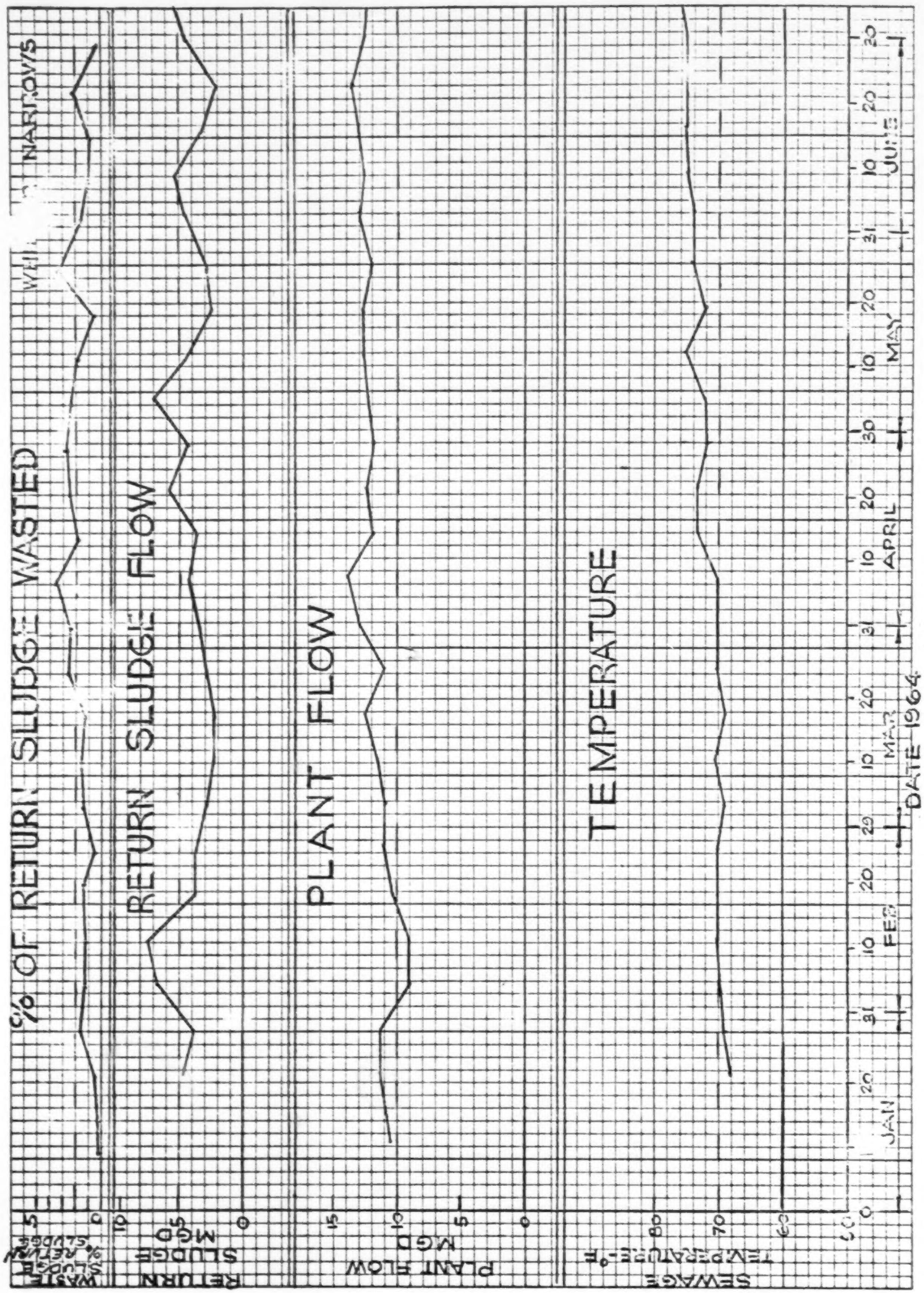


FIGURE Nº 4

FIGURE Nº 4

CLEARPRINT PAPER CO. NO. C3180. 10 DIVISIONS PER INCH BOTH WAYS 70 X 100 DIVISIONS. PRINTED IN U.S.A. ON CLEARPRINT TECHNICAL PAPER NO. 1015

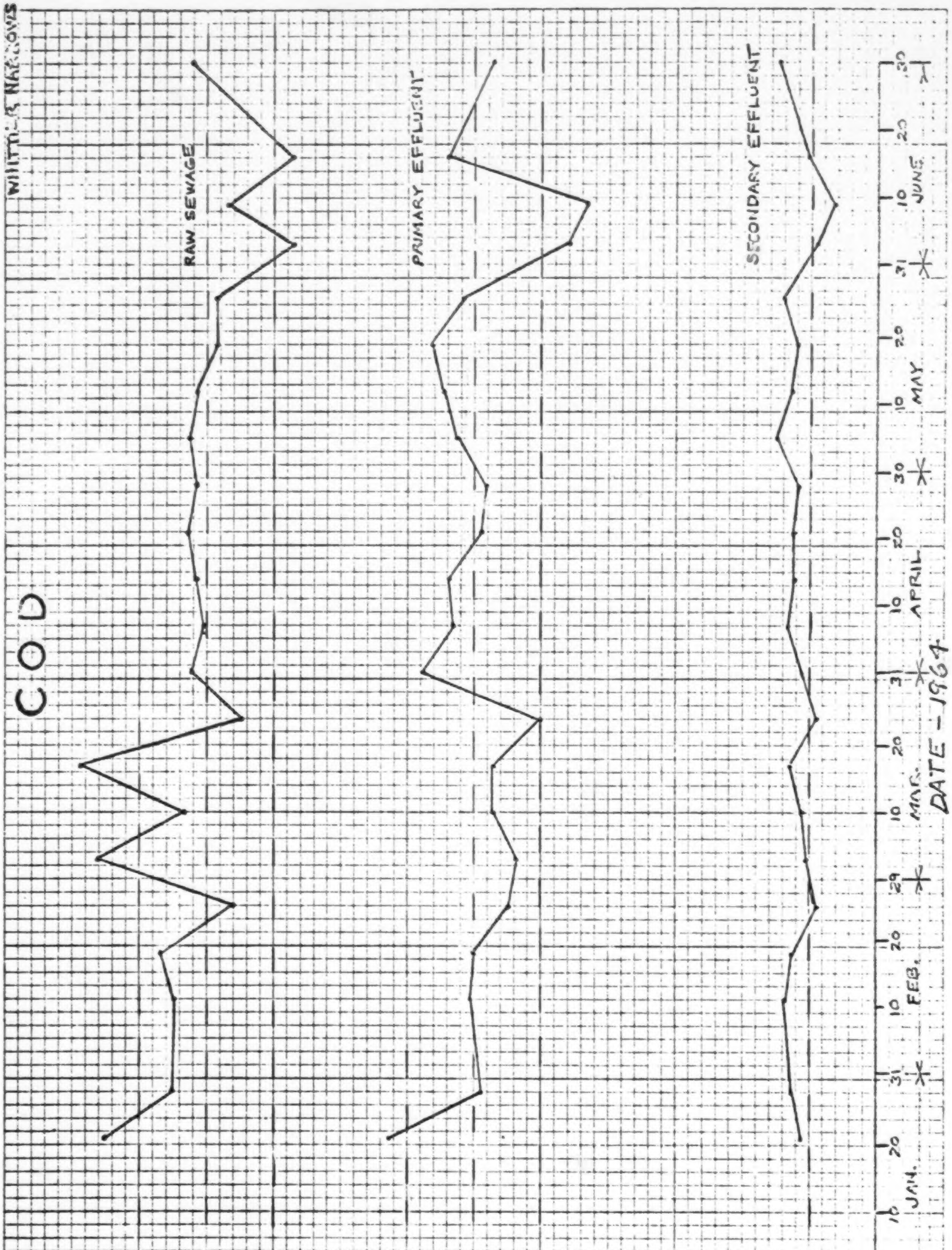


FIGURE N-5

CLEARPRINT PAPER CO. NO. C3190. 10 DIVISIONS PER INCH BOTH WAYS 70 X 100 DIVISIONS PRINTED IN U.S.A. ON CLEARPRINT TECHNICAL PAPER NO. 1015

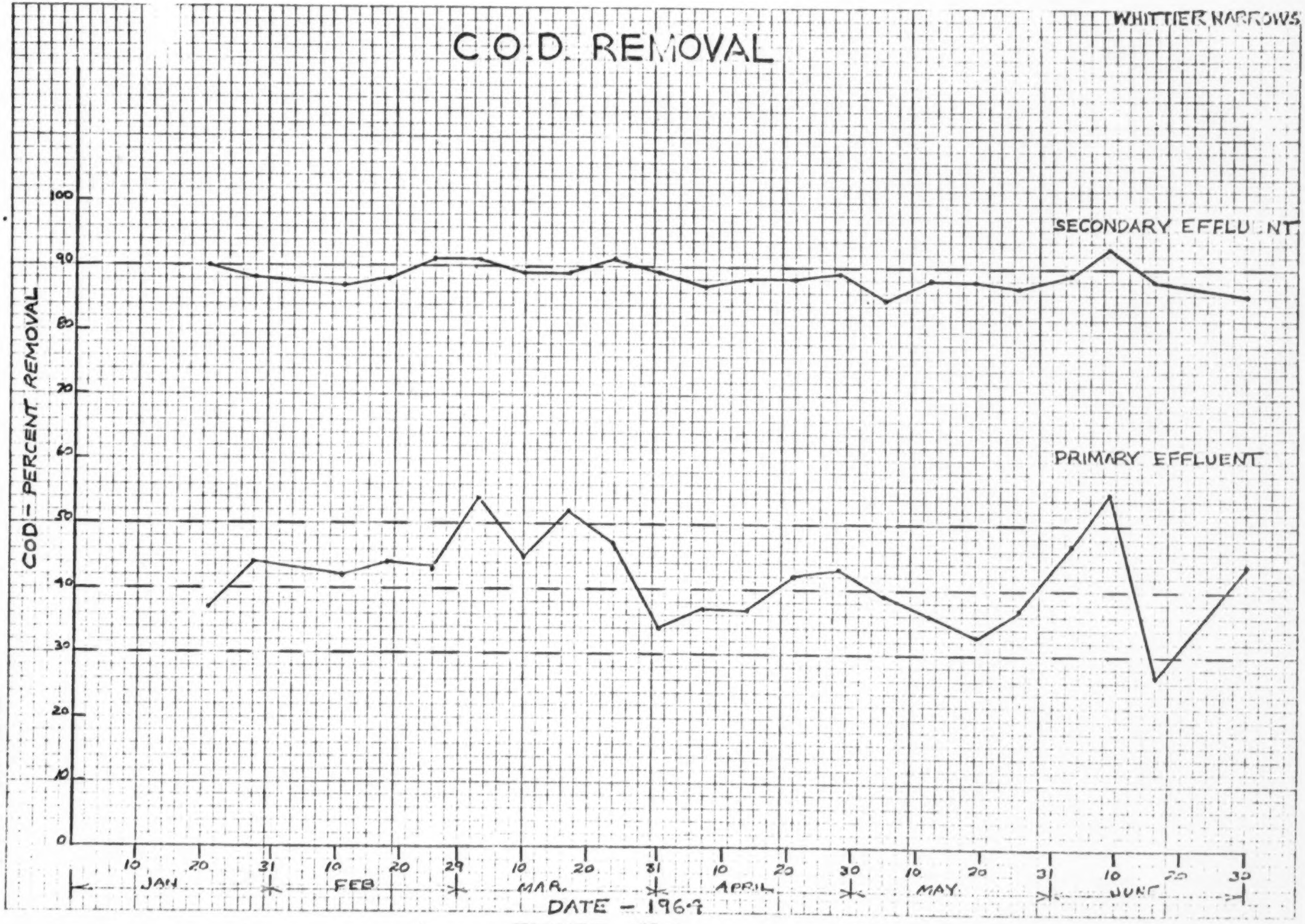


FIGURE Nº 6

FIGURE Nº 6

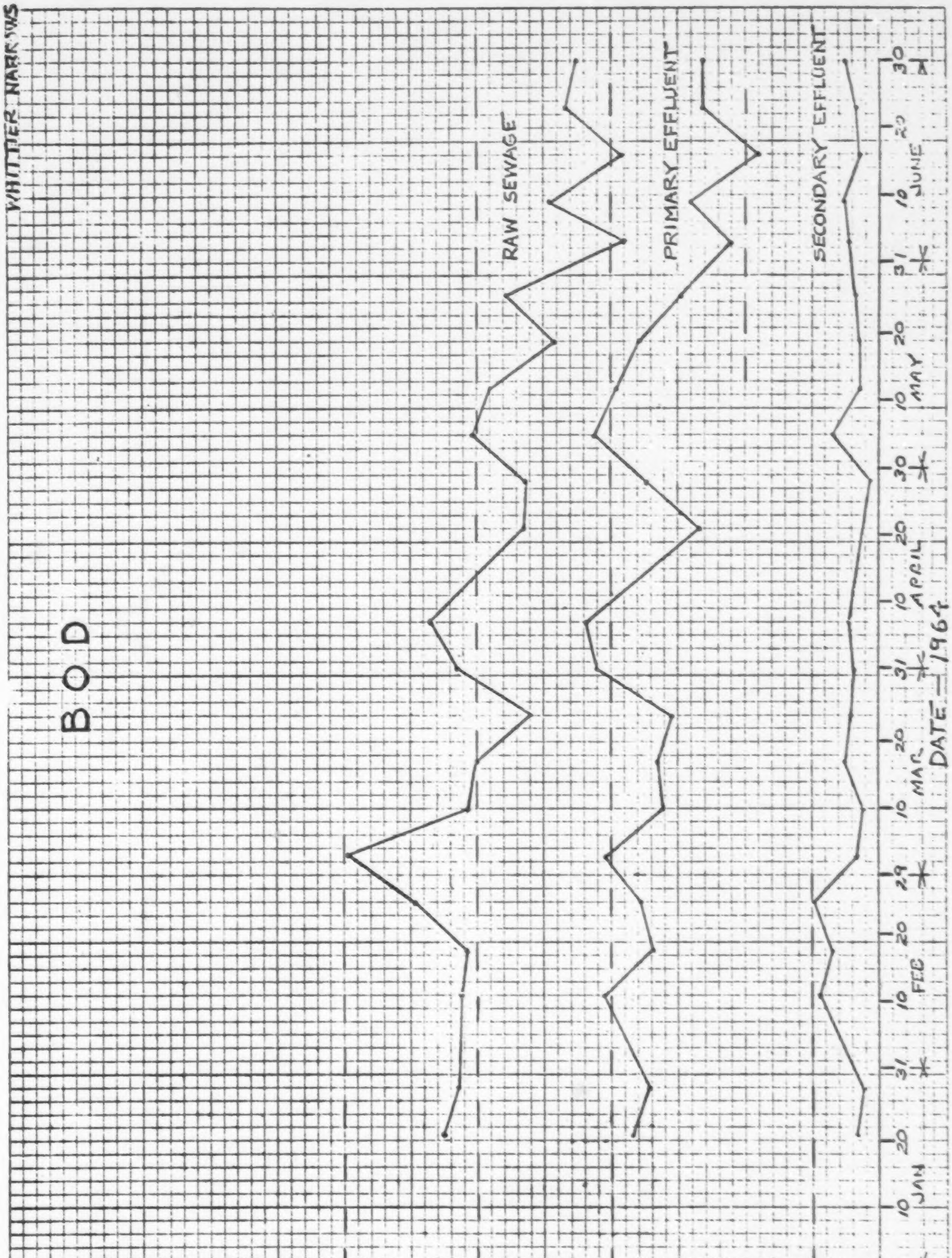


FIGURE N°7

FIGURE N°7

CLEARPRINT PAPER CO. NO. C3180 10 DIVISIONS PER INCH BOTH WAYS 70 X 100 DIVISIONS. PRINTED IN U.S.A. ON CLEARPRINT TECHNICAL PAPER NO. 1015

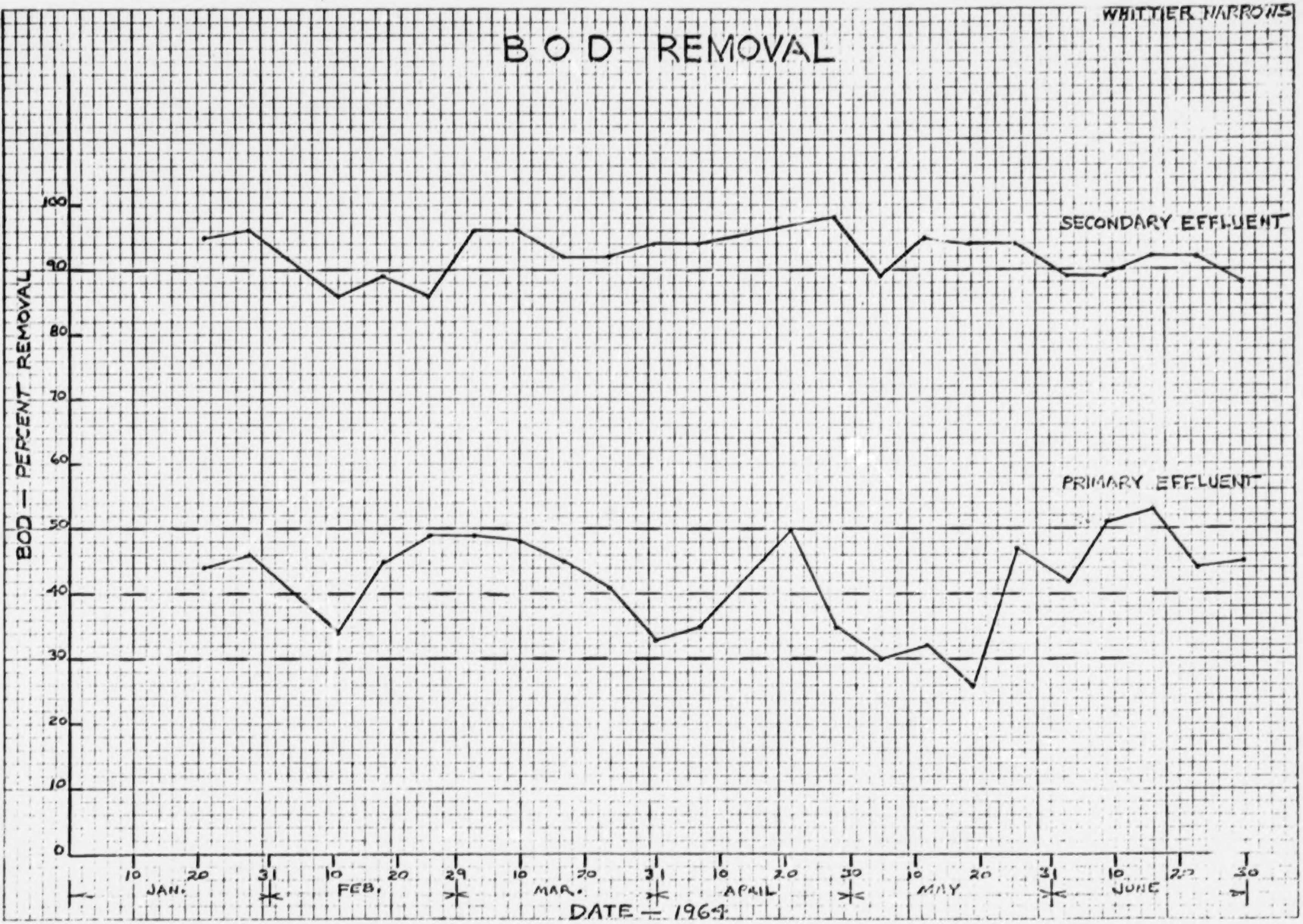


FIGURE N° 8

FIGURE N° 8

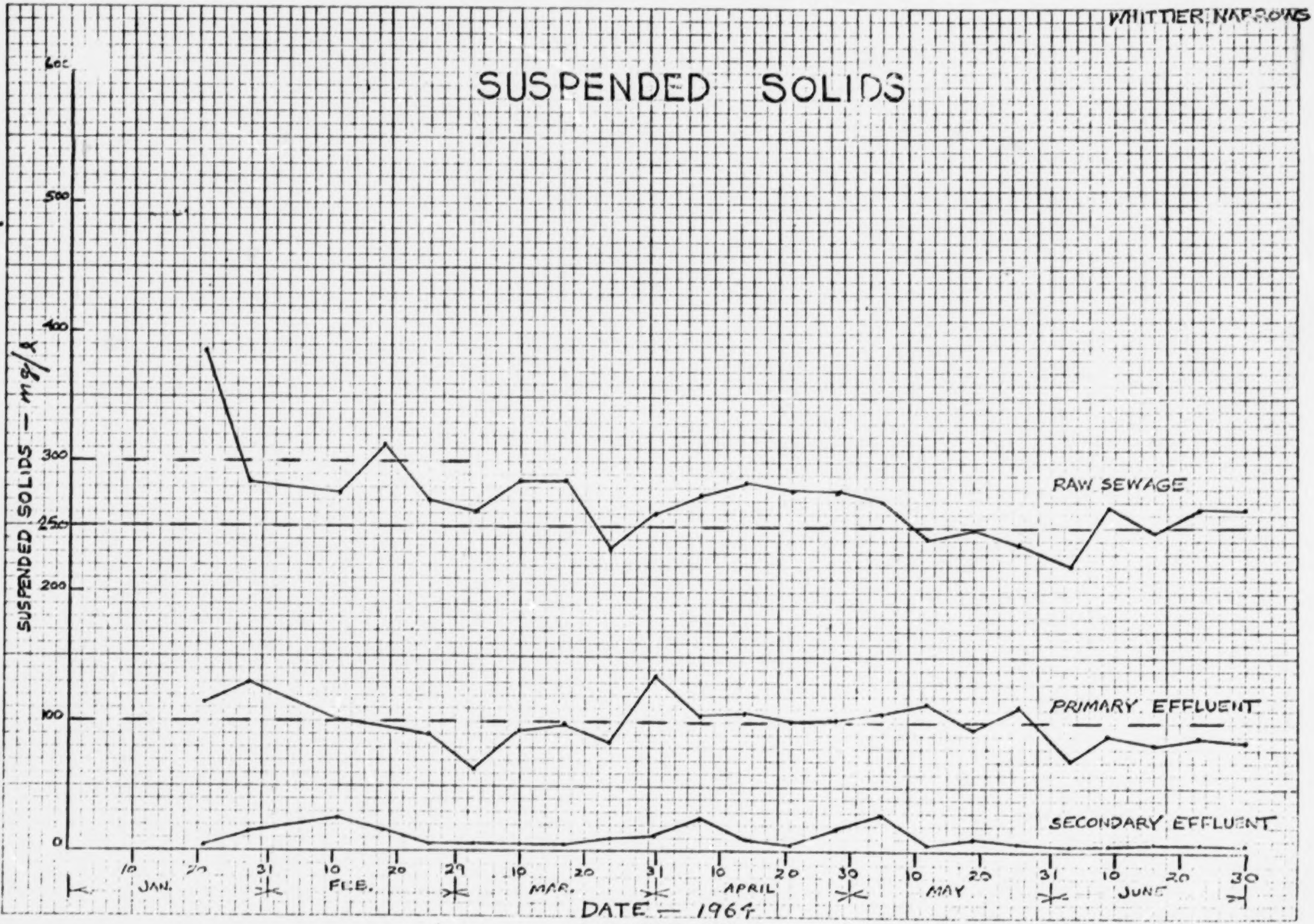


FIGURE N° 9

FIGURE N° 9

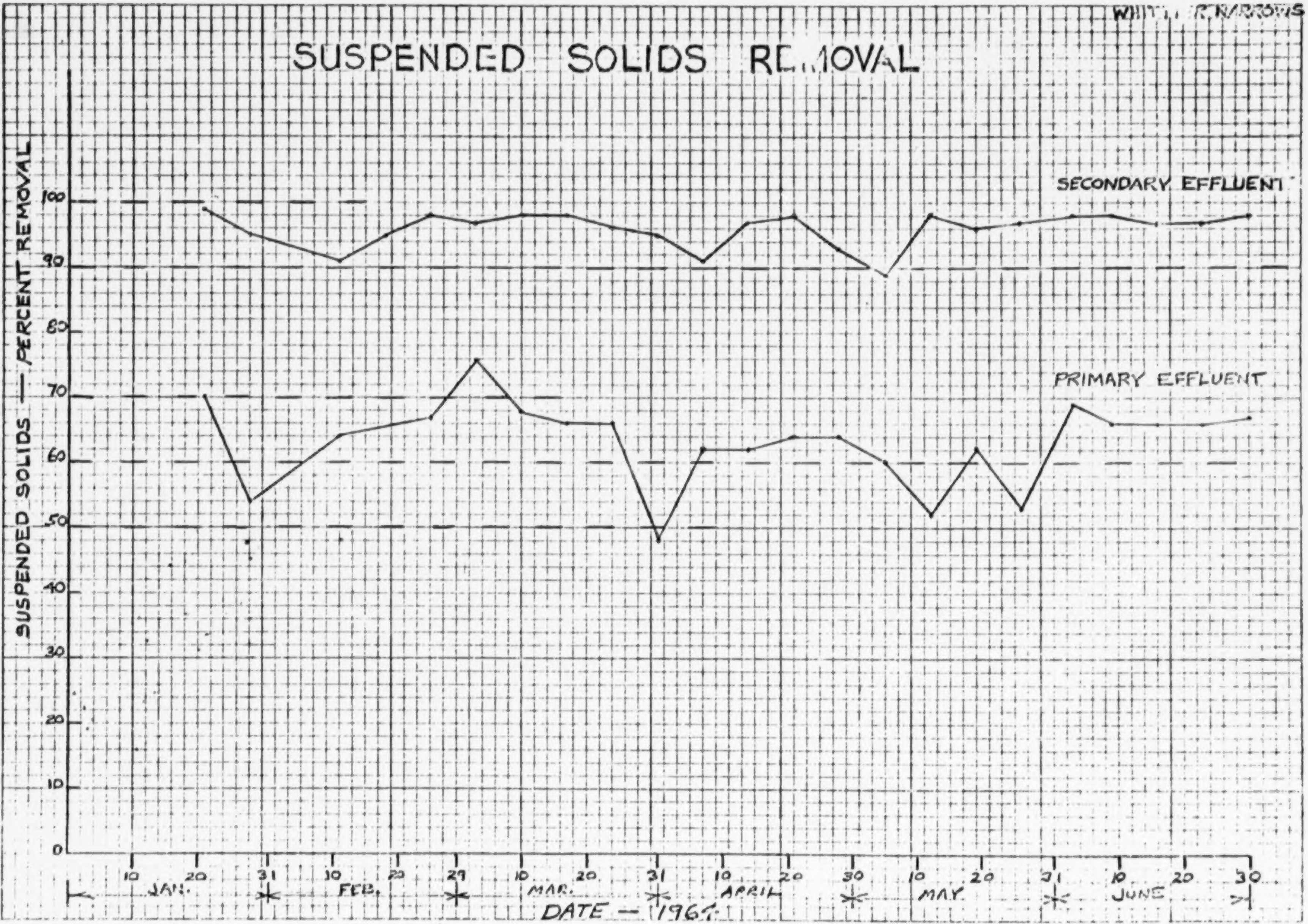

SUSPENDED SOLIDS RE.IOVAL
WHITT... R. NARROWS
SECONDARY EFFLUENT
PRIMARY EFFLUENT
SUSPENDED SOLIDS — PERCENT REMOVAL
100
90
80
70
60
50
40
30
20
10
0
10
20
31
10
20
29
10
20
31
10
20
30
10
20
31
10
20
30
JAN.
FEB.
MAR.
APRIL
MAY
JUNE
DATE — 1964.

FIGURE N° 10

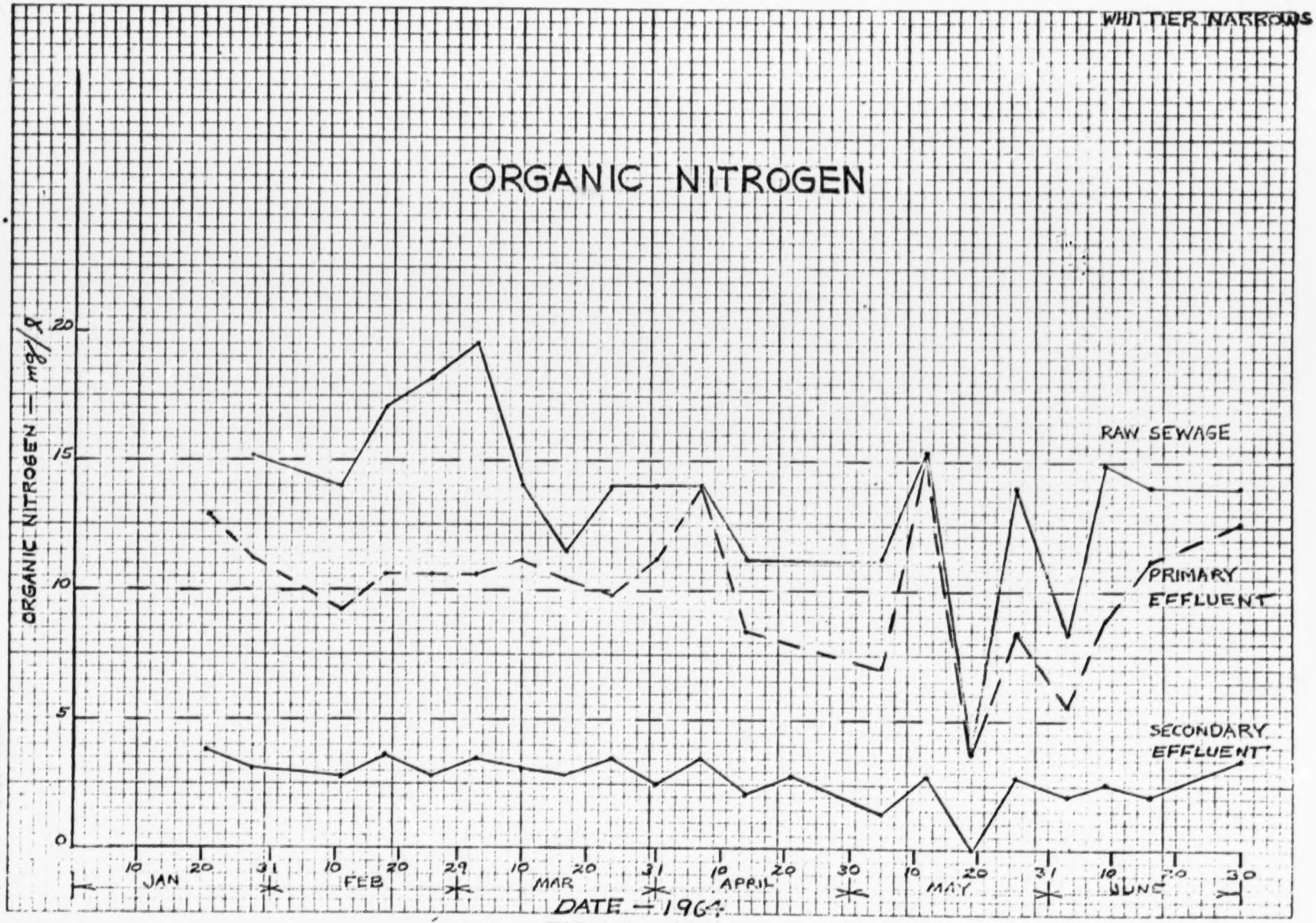
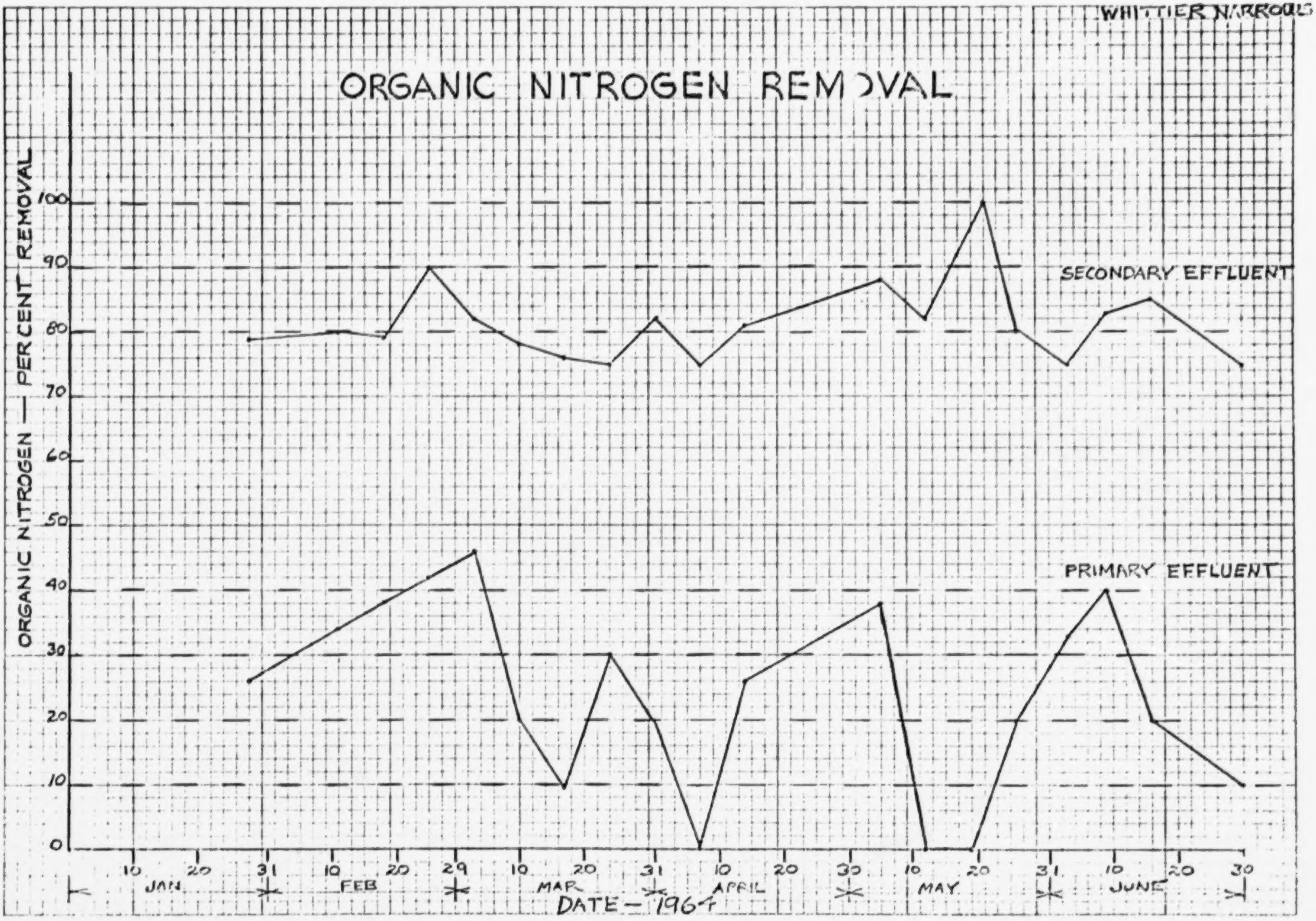


FIGURE N° 11

FIGURE N° 11



ORGANIC NITROGEN REMOVAL

FIGURE No 12

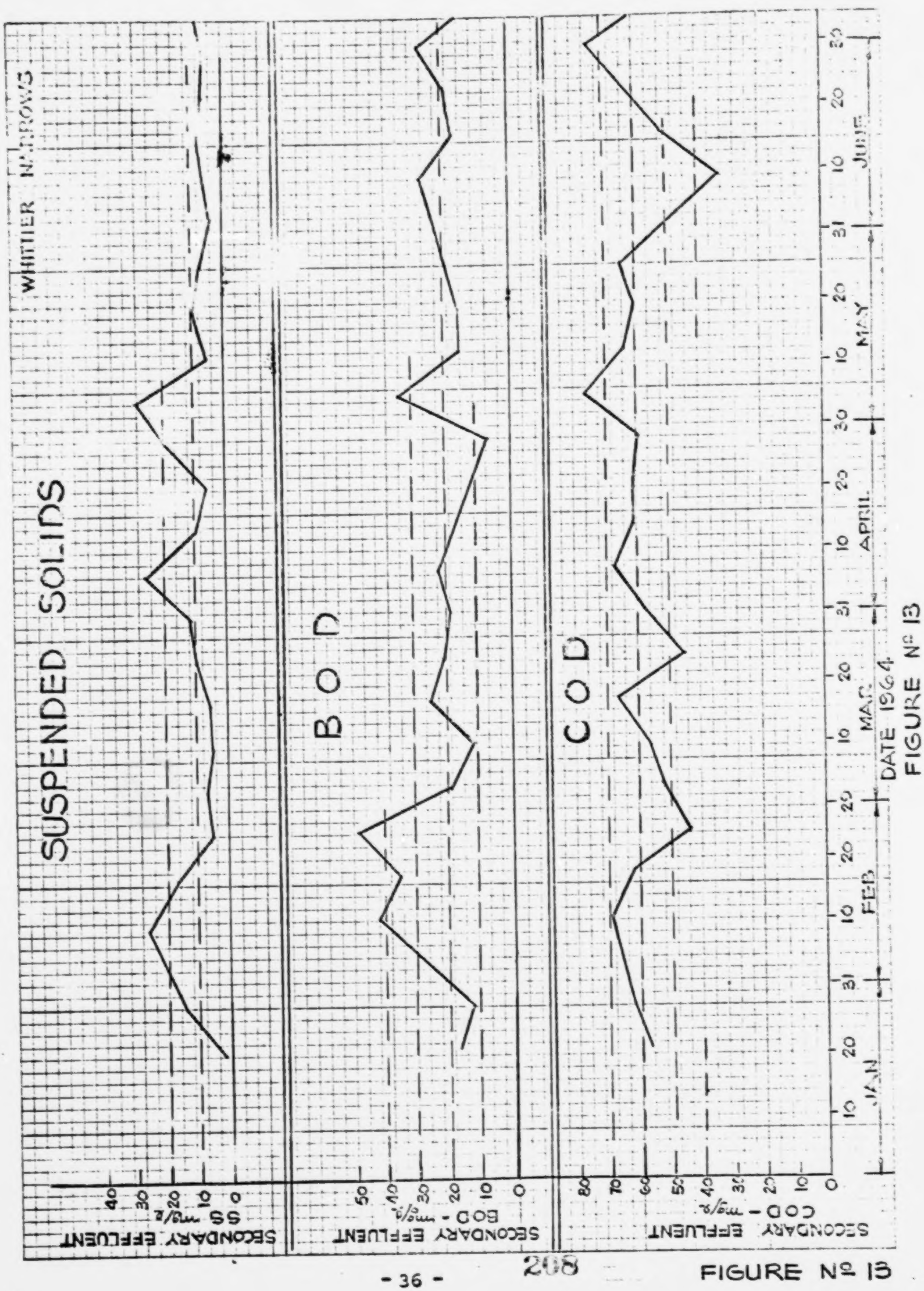

WHITTIER NARROWS
SUSPENDED SOLIDS
B O D
C O D
SECONDARY EFFLUENT SS mg/l
40
30
20
10
0
SECONDARY EFFLUENT BOD - mg/l
50
40
30
20
10
0
SECONDARY EFFLUENT COD - mg/l
80
70
60
50
40
30
20
10
0
10
20
JAN
31
10
FEB
20
29
10
MAR
20
31
10
APRIL
20
30
10
MAY
20
31
10
JUNE
20
30
DATE 1964
FIGURE Nº 13

FIGURE Nº 13

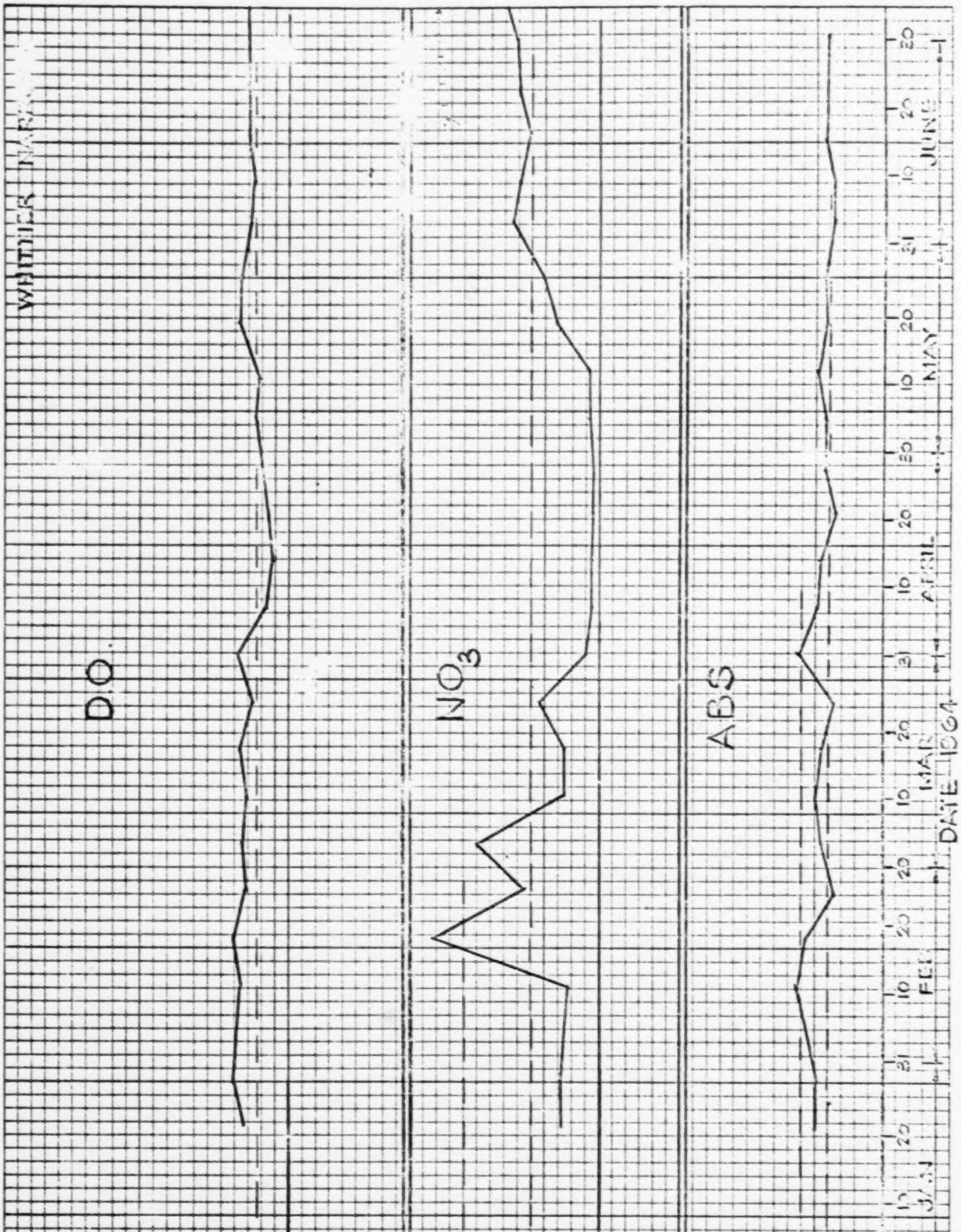


FIGURE Nº 14

FIGURE Nº 14

FINAL EFFLUENT

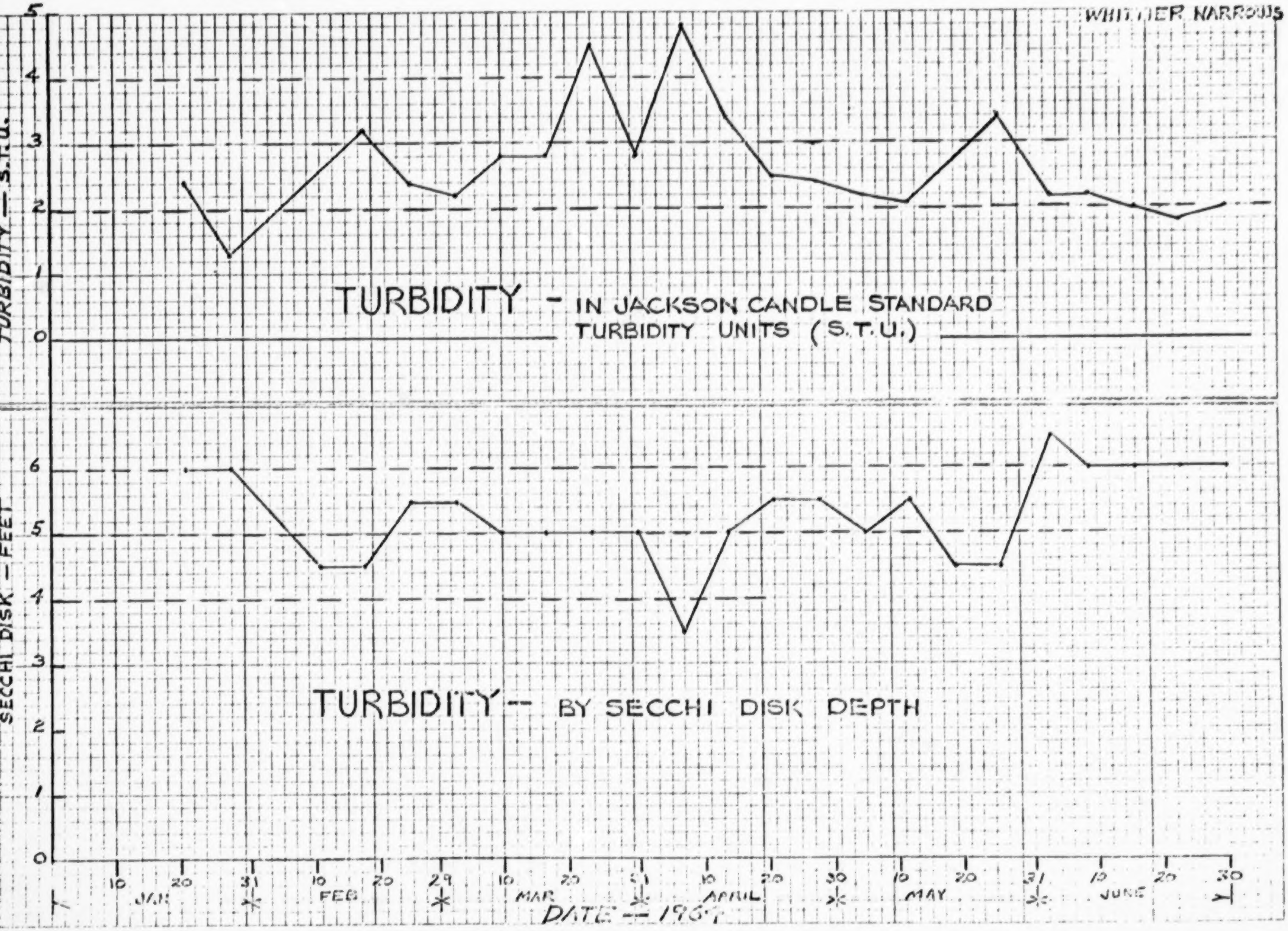


FIGURE Nº 15

210

FIGURE Nº 15

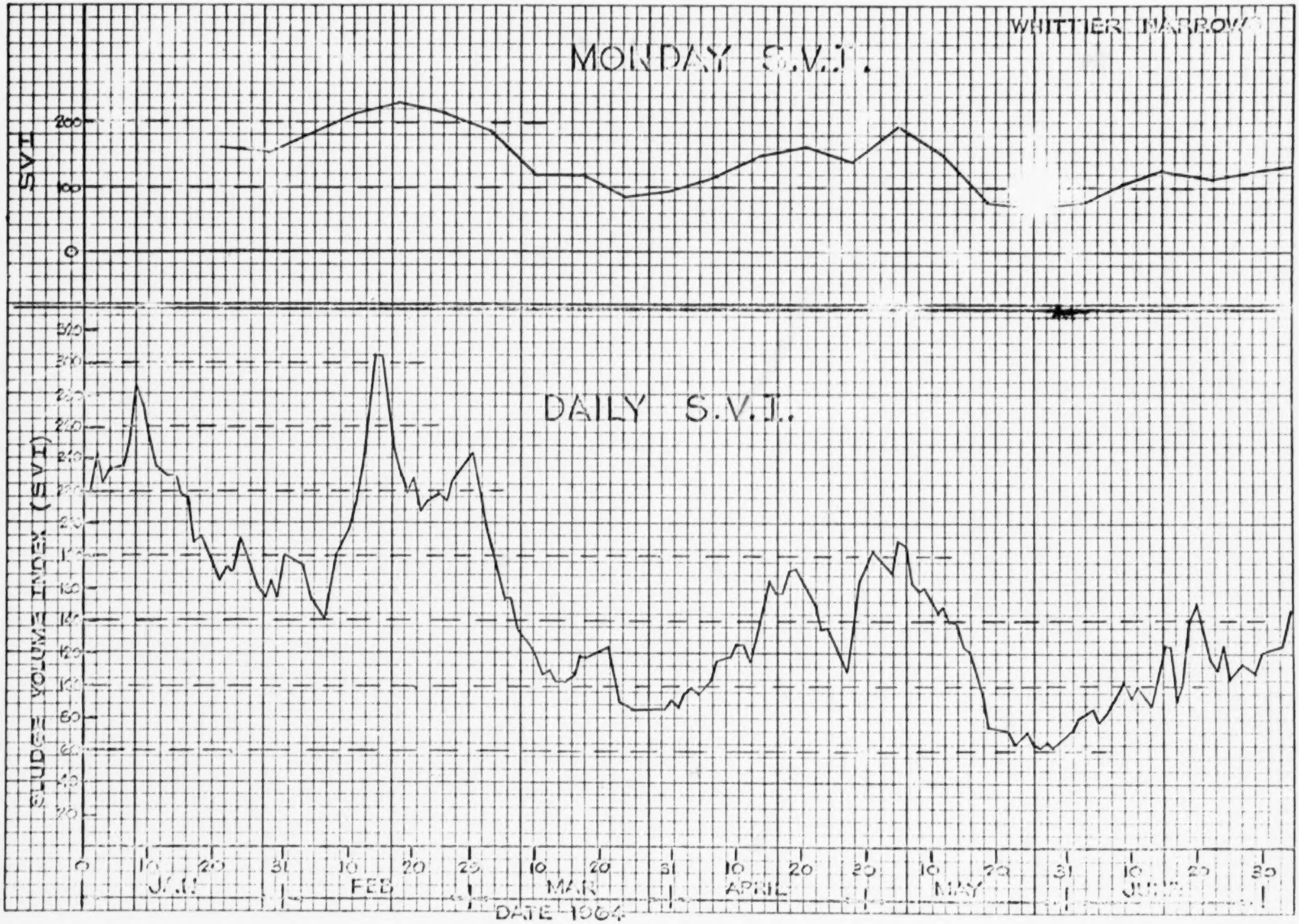

WHITTIER NARROWS
MONDAY S.V.I.
SVI
200
100
0
DAILY S.V.I.
SLUDGE VOLUME INDEX (SVI)
320
300
280
260
240
220
200
180
160
140
120
100
80
60
40
20
0
10
20
31
10
20
29
10
20
31
10
20
30
10
20
31
10
20
30
JAN
FEB
MAR
APRIL
MAY
JUNE
DATE 1964

FIGURE N°16

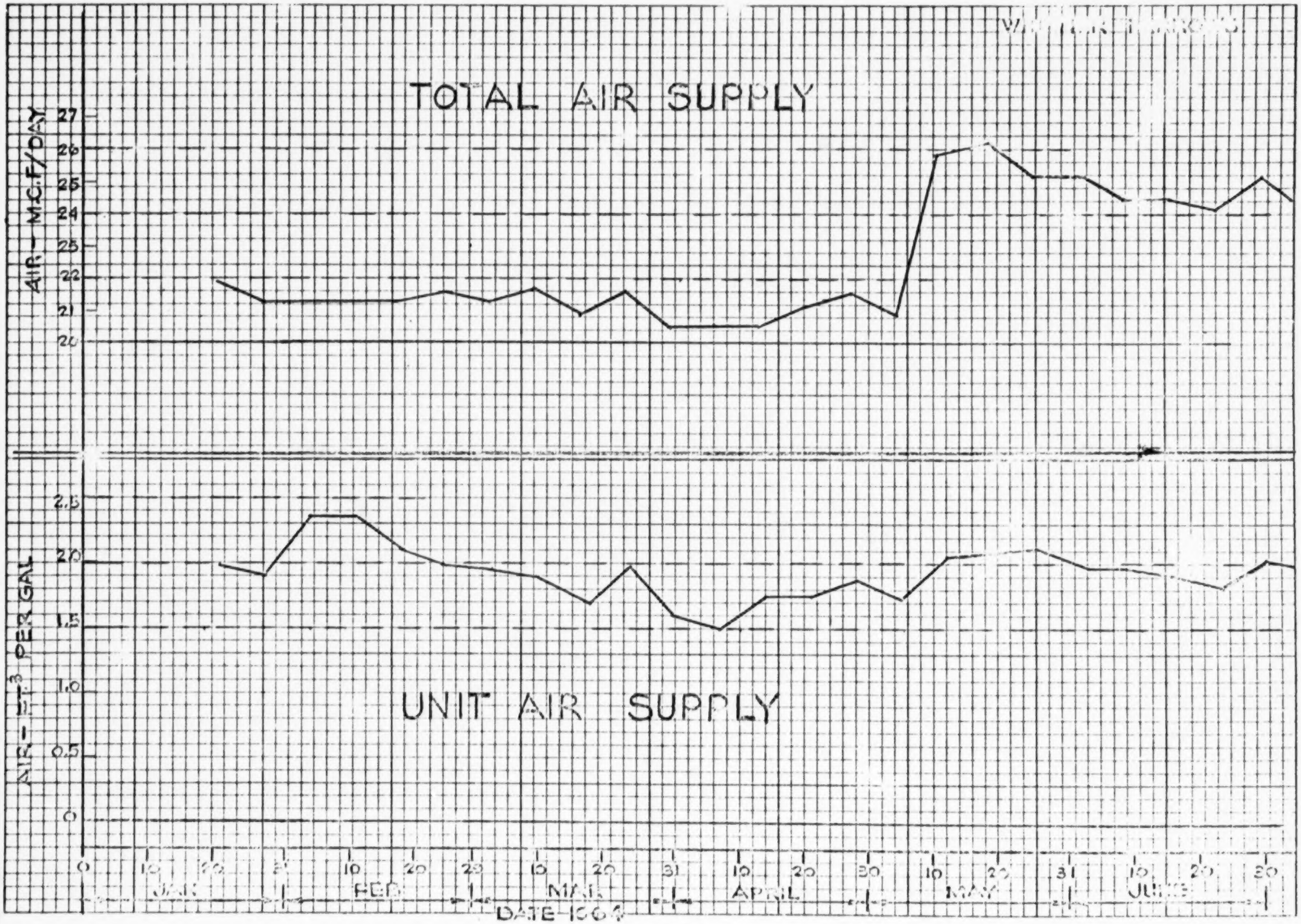

TOTAL AIR SUPPLY
AIR – M.C.F./DAY
27
26
25
24
23
22
21
20
UNIT AIR SUPPLY
AIR – FT³ PER GAL
2.5
2.0
1.5
1.0
0.5
0
0
10
20
31
10
20
29
10
20
31
10
20
30
10
20
31
10
20
30
JAN
FEB
MAR
APRIL
MAY
JUNE
DATE 1964

FIGURE Nº 17

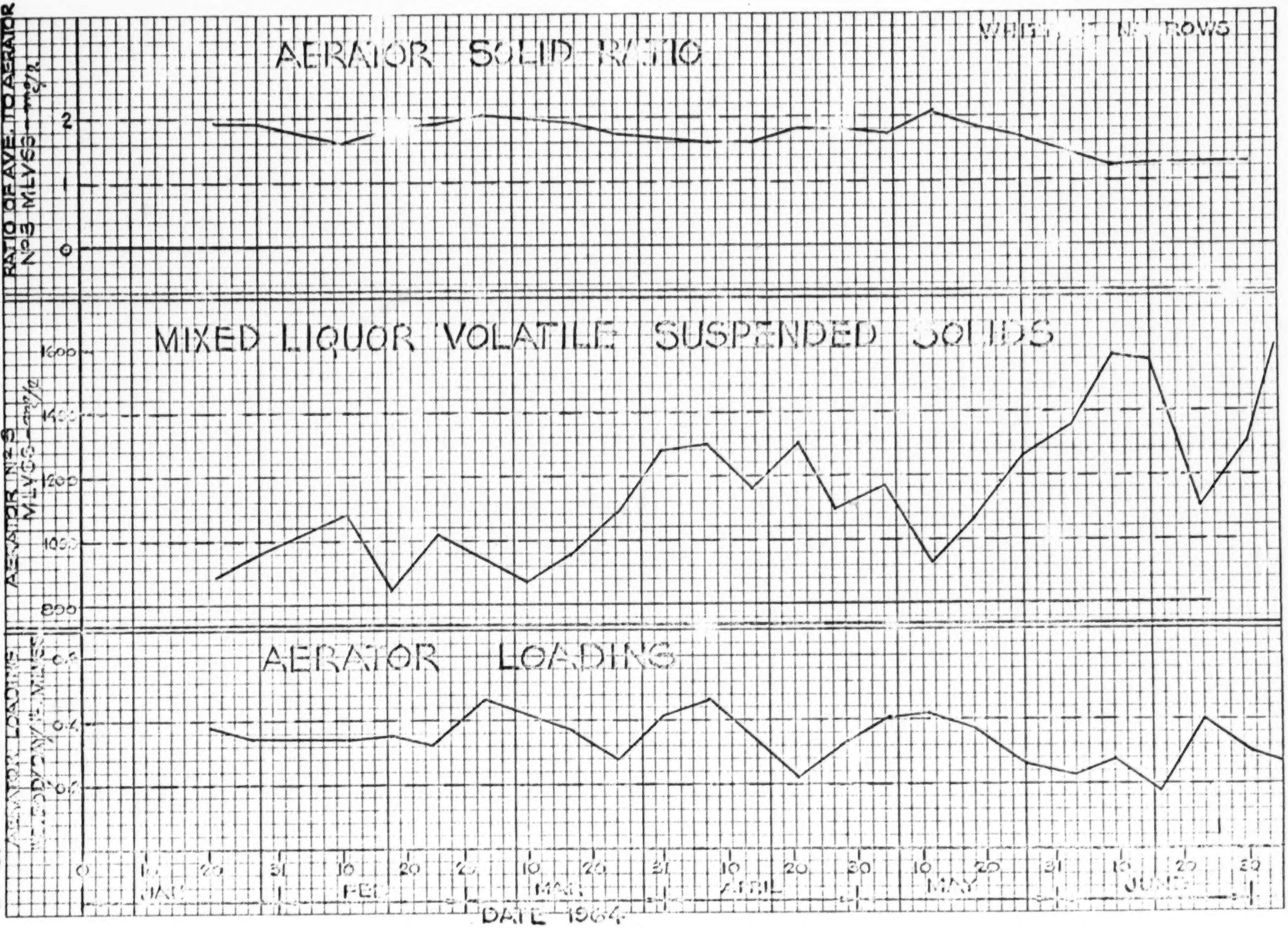

AERATOR SOLID RATIO
RATIO OF AVE. TO AERATOR N°3 MLVSS — mg/l
2
1
0
MIXED LIQUOR VOLATILE SUSPENDED SOLIDS
AERATOR N°3 MLVSS — mg/l
1600
1400
1200
1000
800
AERATOR LOADING
0.6
0.4
0.2
0
10
20
31
JAN
10
20
29
FEB
10
20
31
MAR
10
20
30
APRIL
10
20
31
MAY
10
20
30
JUNE
DATE 1964

FIGURE N°18

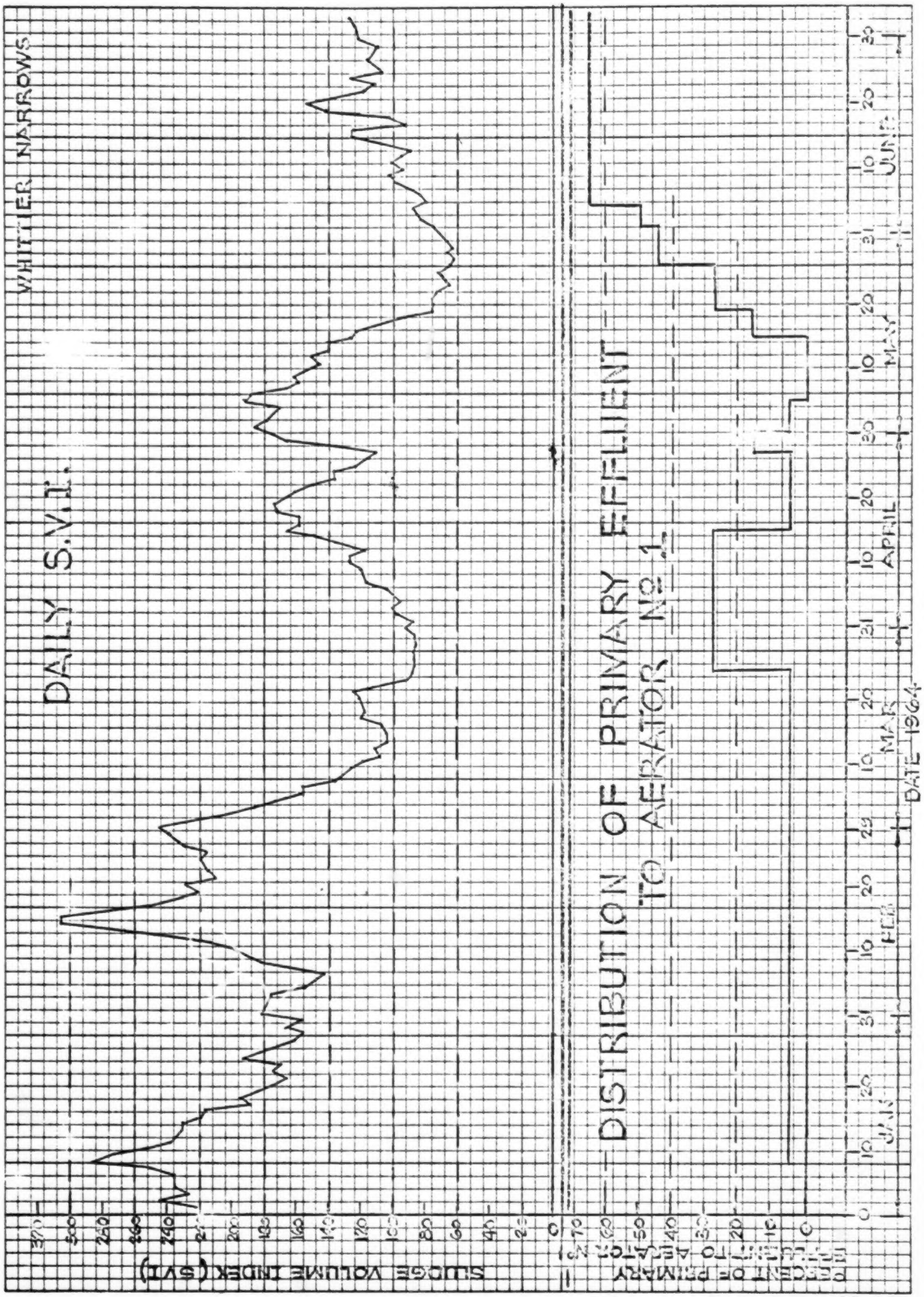


FIGURE N° 19

FIGURE N° 19

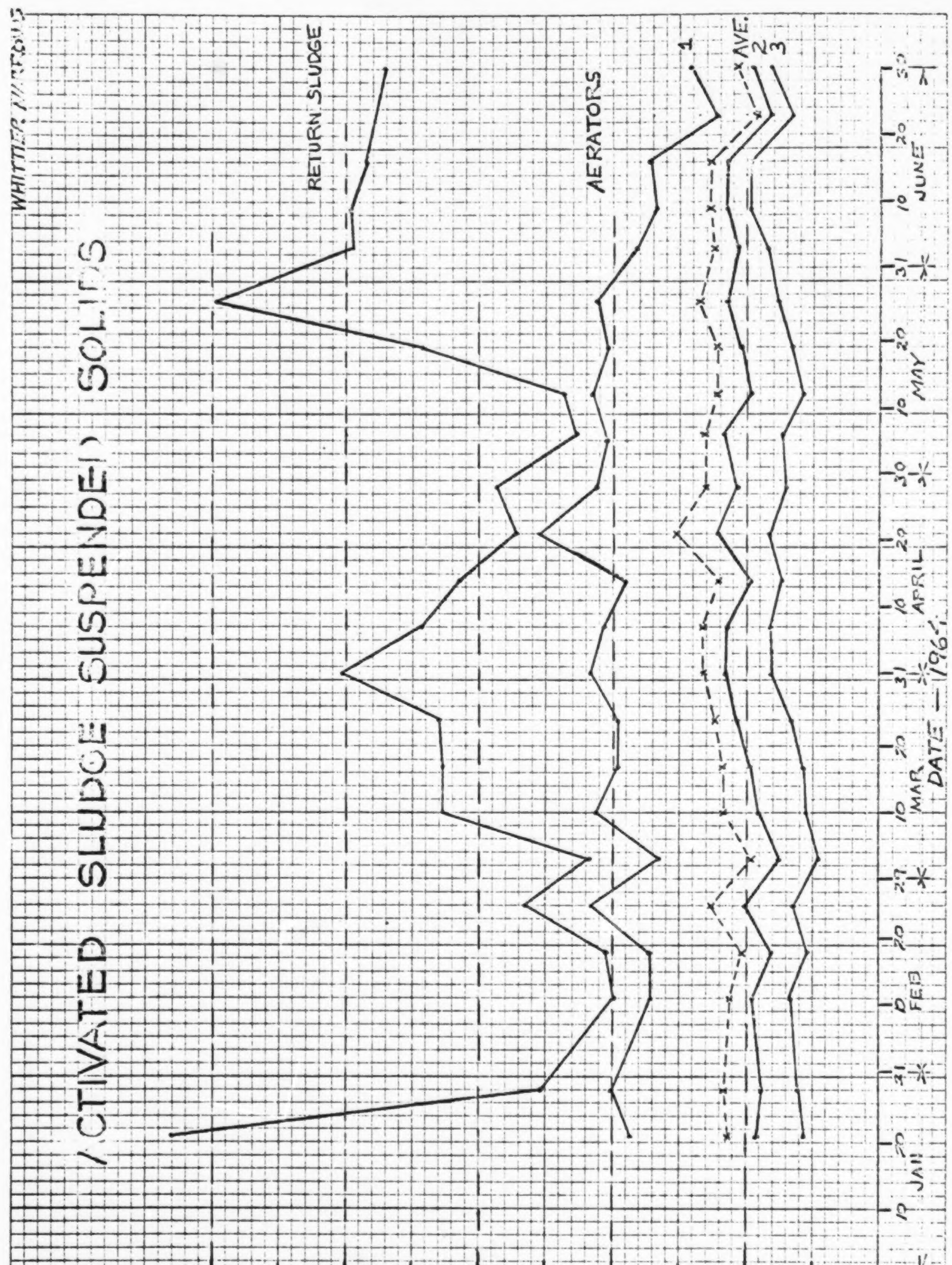
ACTIVATED SLUDGE SUSPENDED SOLIDS
WHITTIER NARROWS
RETURN SLUDGE
AERATORS
1
AVE.
2
3
SUSPENDED SOLIDS — mg/l
1000
2000
3000
4000
5000
6000
7000
8000
9000
10,000
11,000
12,000
10
20
JAN
31
10
FEB
20
29
10
MAR
20
31
10
APRIL
20
30
10
MAY
20
31
10
JUNE
20
30
DATE — 1961
FIGURE 20

FIGURE N° 20

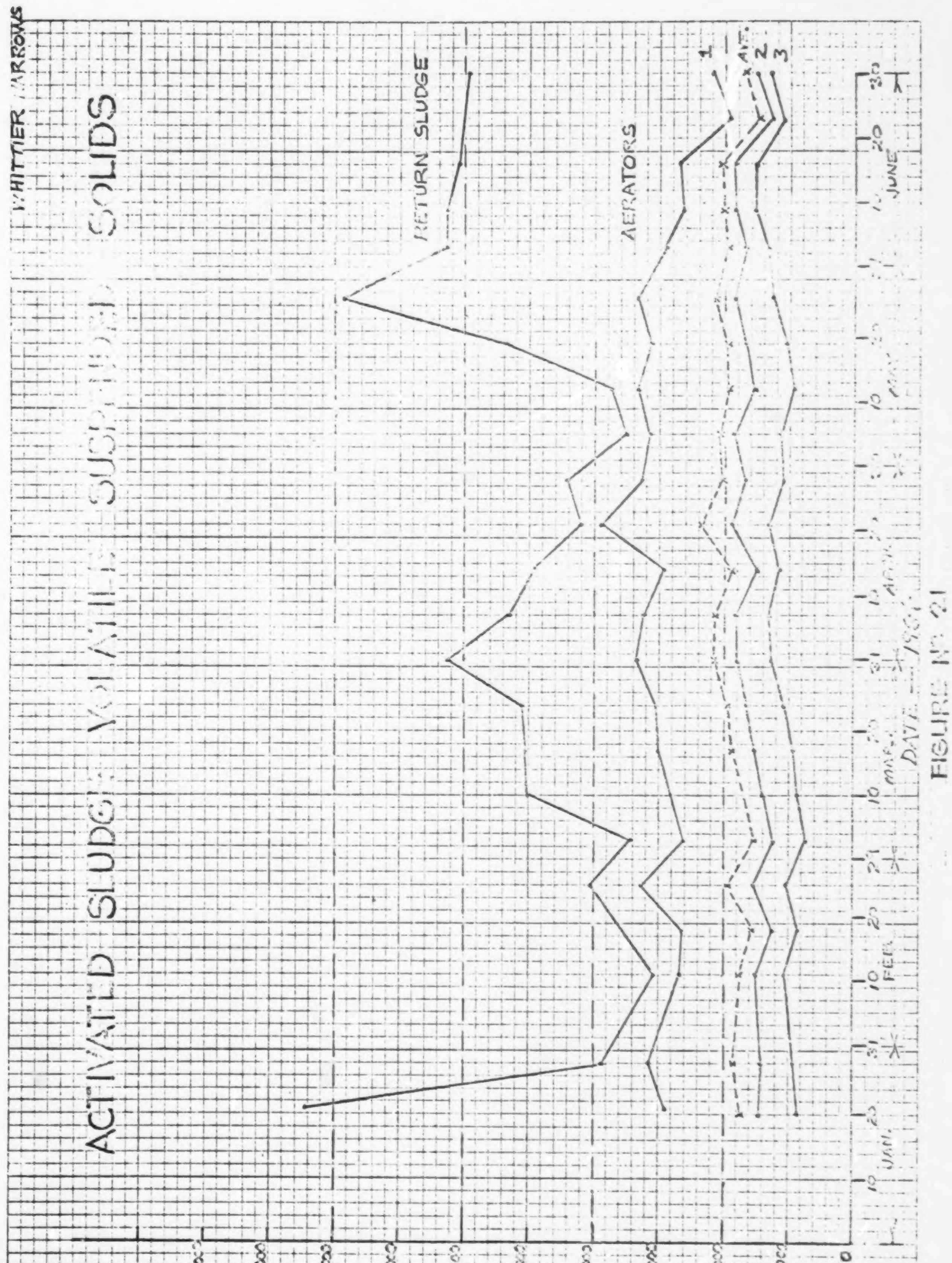
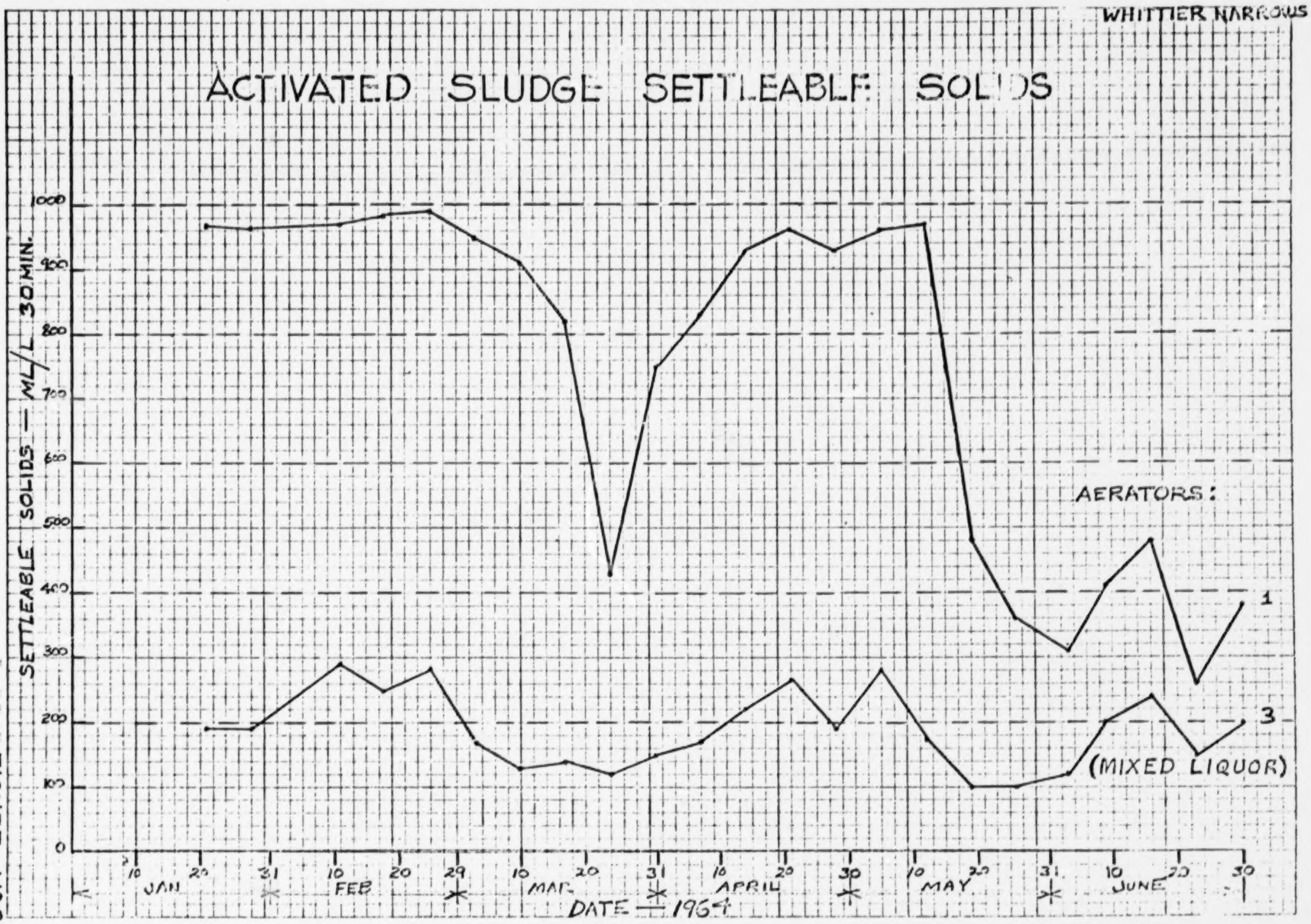


FIGURE Nº 21

FIGURE Nº 21



WHITTIER NARROWS
ACTIVATED SLUDGE SETTLEABLE SOLIDS
SETTLEABLE SOLIDS — ML/L 30 MIN.
1000
900
800
700
600
500
400
300
200
100
0
AERATORS:
1
3
(MIXED LIQUOR)
10
20
31
10
20
29
10
20
31
10
20
30
10
20
31
10
20
30
JAN
FEB
MAR.
APRIL
MAY
JUNE
DATE — 1964

FIGURE N° 22

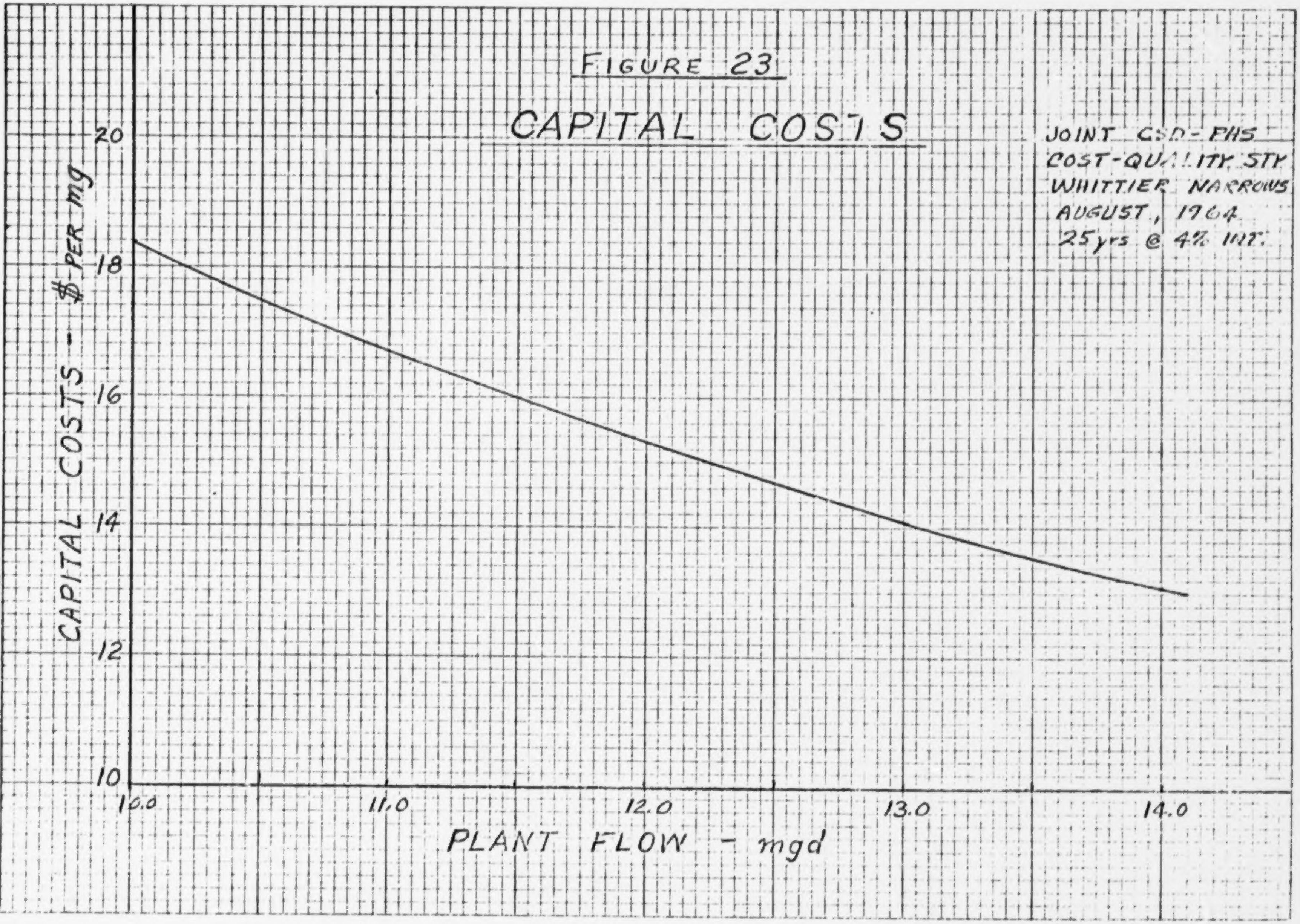


FIGURE 23

CAPITAL COSTS

JOINT CSD-PHS
COST-QUALITY STY
WHITTIER NARROWS
AUGUST, 1964
25yrs @ 4% INT.

FIG. 23

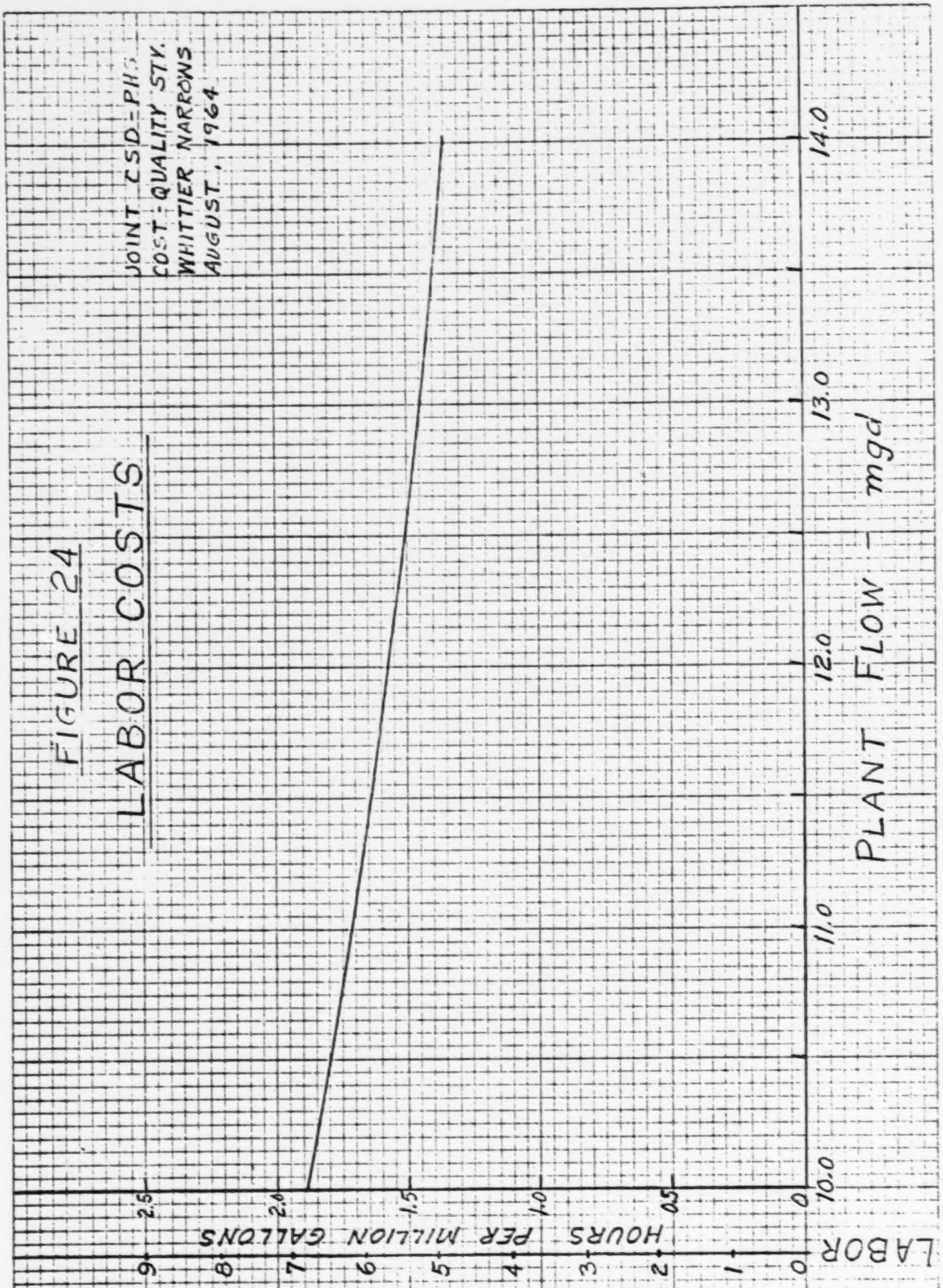


FIGURE 24

LABOR COSTS

JOINT CSD-PHS
COST: QUALITY STY.
WHITTIER NARROWS
AUGUST, 1964

FIG. 24

TABLE 1

CAPITAL COSTS FOR THE WHITTIER NARROWS WATER RECLAMATION PLANT

Unit	Capital Cost	
Two Primary Sedimentation Tanks	$ 235,000	
Three Aeration Tanks	480,000	
Five Final Sedimentation Tanks	250,000	
Blowers	80,000	
Sub-Total (For units related to effluent quality)		$1,045,000
Control Building	265,000	
Pumps	35,000	
Chlorine Contact Tanks & Equipment	45,000	
Preliminary Sitework & Misc. Structures	55,000	
Sub-Total		$ 400,000
TOTAL		$1,445,000

Amortization of $1,045,000 over 25 years at 4% interest = $67,000 per year.

All capital costs include 7% for engineering.

TABLE 2

COMPARISON OF PLANT OPERATION WITH EITHER ONE OR TWO PRIMARY SEDIMENTATION TANKS

Units	No. In Use	Detention Time (Hrs.)	GPD Per Ft. Of Weir X 10^3	GPD Per Ft^2	Lbs. BOD Per Day Per lb. MLVSS	Effluent Characteristics							
						COD		BOD		S.S.		NO_3-N	
						mg/1	% Removal	mg/1	% Removal	mg/1	% Removal	mg/1	% of T.N.
Primary[1]	2	2.4	52	900	-	307	42	181	44	109	64	-	-
Aerators	3	1.9	-	-	0.355	-	-	-	-	-	-	-	-
Final	5	1.6	12	1110	-	60	89	31	99	13	96	5.2	15
Primary[2]	1	1.0	117	2100	-	293	42	162	42	96	63	-	-
Aerators	3	2.7	-	-	0.348	-	-	-	-	-	-	-	-
Final	5	1.6	13	1087	-	57	89	19	93	10	96	2.8	9

[1] Two primary tanks were used from March to June.

[2] One primary tank was used from January to February.